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A THEORETICAL AND EXPERIMENTAL APPROACH TO AN  
ANAEROBIC-AEROBIC WASTE-WATER TREATMENT FACILITY

A THESIS

Presented to

The Faculty of the Graduate Division

by

Woodrow Wilson Albury, Jr.

In Partial Fulfillment

of the Requirements for the Degree

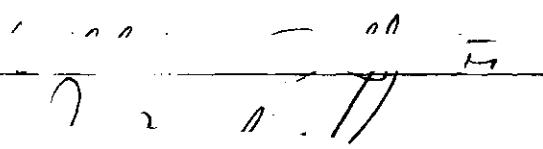
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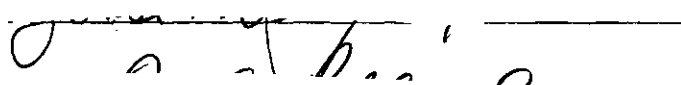
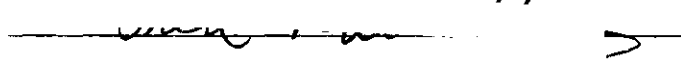
Georgia Institute of Technology

September, 1967

A THEORETICAL AND EXPERIMENTAL APPROACH TO AN  
ANAEROBIC-AEROBIC WASTE-WATER TREATMENT FACILITY

Approved:

  
Chairman

  
  
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## SUMMARY

The theoretical, practical, and economical aspects of the anaerobic treatment and the aerobic treatment processes were investigated. The purposes of this research were (1) to develop theoretical process performance equations for the anaerobic and the aerobic processes, (2) to develop a method for an economic comparison of the anaerobic-aerobic process with the aerobic process, and (3) to construct and operate a laboratory scale, anaerobic-aerobic reactor in an attempt to substantiate the theoretical development.

The process control constants and parameters as reported in the literature were utilized for the theoretical development of the process performance equations for both the anaerobic and the aerobic systems. The equations were evaluated and graphs developed so that the effect on the system of variations of the controlling parameters is readily apparent. These equations were then used in the economic evaluation of the two systems.

The laboratory scale, anaerobic-aerobic reactor was constructed and operated for a period of six months. The reactor was fed continuously at the rate of 50 liters per day at an influent COD of 1000 mg/L COD. A two-day detention time through the unit was used.

The results from the theoretical evaluation clearly indicate that process performance can be easily predicted. The economic evaluation proved that for the selected parameters, an anaerobic-aerobic treatment unit will result in an annual savings of \$13,000.00

over that of an aerobic unit. The laboratory investigations proved that an anaerobic-aerobic treatment unit is feasible.

Further investigations to obtain additional data to substantiate the theoretical development is strongly suggested.

## CHAPTER I

### INTRODUCTION

The objective of the present method of waste-water treatment is to protect man from his own wastes and to minimize the impact of his waste on the aquatic ecosystem of the streams, rivers, and lakes. This aquatic ecosystem is a relatively ordered biological community which, through action and reaction among its various components, maintains itself at steady-state--provided no external stimuli upset the steady-state conditions.

The waste-waters produced by man, both domestic and industrial, are available sources for this external stimuli, and have continuously been discharged into the aquatic environment. The effect has been a continual degradation of water quality throughout the nation. Streams and lakes, polluted to the extent that they are unfit for any use except open sewers, are now common.

As a result of man's supplying the stimuli to the aquatic environment, he has necessarily become a part of that environment. However, since his effect on the environment is detrimental to his own existence, he has attempted to isolate himself from the aquatic system by the construction of water treatment plants. For years, these treatment plants successfully purified, for man's consumption, the waters which he polluted. However, as the streams became more and more polluted, the assimilative capacity of the aquatic environment was over-

come and the economics of water treatment became staggering. At this point, man was forced to accept his responsibility in the aquatic environment and he began to treat his waste before discharge to the receiving waters.

Presently, there are two major types of waste-water treatment: aerobic treatment and anaerobic treatment. Each is substantially different from the other but their common objective is removal of pollutional characteristics from the waste discharges.

These pollutional characteristics may be numerous, but the most common measures include: organic content (BOD or COD), suspended solids, dissolved solids, temperature, pH, and coliform count.

Because of the basic nature of the processes, there are advantages and disadvantages to both the aerobic and the anaerobic systems. The various modifications of the aerobic process include activated sludge, contact stabilization, dispersed growth aeration, step aeration, and extended aeration. All involve the mixing of the organics in the waste-water with a bacterial population in a favorable environment. When supplied with adequate oxygen, the bacteria remove the organic matter and utilize it, along with the supplied oxygen, to produce more bacteria and to maintain cell metabolism. Because of the high energy yield of the aerobic oxidation process, only a small portion of the removed organic matter is used to maintain cell metabolism while the larger portion is converted to cell structure. Therefore, only the smaller portion is truly stabilized and the principal result is the concentrating of the organic matter in the bacterial population, which is no more stable than the original waste flow. However, after concentration,

the bacteria may be easily removed from the waste flow and further stabilization accomplished through endogenous respiration.

The advantage to this type of treatment is that rapid growth rates permit relatively short detention times which decrease the initial cost of the facility. The aerobic process is fairly simple to operate and is capable of absorbing shock loads. The largest disadvantages are that the operating costs are high and the theoretical upper limit of the organic concentration for economical treatment is set by the rate at which oxygen can economically be supplied. This upper limit is approximately 800 mg/L BOD<sub>5</sub> (3).

The two basic anaerobic processes are the conventional anaerobic process and the anaerobic contact process. The conventional anaerobic process has been used for years but principally as a secondary treatment process for disposing of accumulated waste sludges from primary sedimentation, activated sludge and trickling filters. Because of the high accumulation of organic solids, low nutrient concentration, low growth rates, lack of mixing and intermittent feeding, detention times of 15 to 40 days were necessary to stabilize the sludge. In effect, this treatment represented little more than providing time for endogenous respiration to take place.

Only in the last 20 years has there been considerable investigation into the use of an anaerobic contact process for treatment of concentrated waste-waters. This type of treatment is much more complex than that of the aerobic system, since successful stabilization of the organic matter is dependent upon a critical balance between the acid-forming bacteria and the methane-forming bacteria (1,2).

In the anaerobic contact process organic matter is oxidized by intramolecular oxidation. This type of oxidation does not produce oxidized inorganics nor does it provide much energy to support the growth of the bacteria (3,4,5). Therefore, only small quantities of the organic matter are converted to cell mass while major quantities are converted to methane and carbon dioxide.

In terms of end products, the anaerobic process is much better suited to the objectives of stabilization than is the aerobic process. However, the biological growth rates for the anaerobic system are much lower than those of the aerobic system. Consequently, for dilute waste-waters, anaerobic treatment is not economically acceptable. At high input concentrations, however, low growth rates can be tolerated and high strength waste-waters can be stabilized by anaerobic sludge digestion. The minimum acceptable input concentration for this approach is reported to be 1 per cent organic solids (3).

At low organic concentrations, aerobic treatment is the most economical while at very high organic loadings, anaerobic treatment must prevail. In the transition region, however, from 800 mg/L BOD<sub>5</sub> to 1 per cent organic solids, there has yet to be established basic alternative flow patterns for waste-water treatment procedures. It is within this transition region that many investigators have attempted to develop a suitable modification of the anaerobic sludge digestion process such that it could be used as an alternative to or in conjunction with the aerobic process in treating all the waste-water and not just the concentrated aerobic biomass. The resulting anaerobic processes have been generally referred to as anaerobic contact processes. The absence of

such a process which is generally applicable prevents an economical and technological optimization of the treatment of waste-waters, usually industrial process waste-waters, falling in this concentration region.

#### Comments on Previous Work

In recent years, there have been numerous investigations into the applicability of the anaerobic contact process for treating various types of concentrated organic waste-waters. Most of these investigations have been batch or semi-continuous flow in nature and all have indicated that the anaerobic process can successfully treat different types of waste water. There is no point in discussing the various statistics of each investigation. The basic conclusion is that the process is feasible but the major requirement for successful operation is an efficient solids separation and recycle device and a method for predicting process performance. The anaerobic process has been successfully applied to the treatment of starch-gluten wastes (6), citrus waste (7), distilling waste (8), tomato and pumpkin waste (9), packing house waste (25,11), and synthetic milk waste (3).

Some of the most advanced investigations of the anaerobic contact process have been conducted by Schroepfer et al. (10), and Schroepfer and Ziemke (11). The investigations have shown that the anaerobic contact process can successfully treat packing house waste, synthetic milk waste, fatty acid waste, and wood fiber waste. BOD reductions of 70 per cent to 97 per cent have been reported at detention times as low as 0.25 days. The effects on system performance of various process parameters is also discussed. These parameters include

the unit loading rate, pH, solids concentration, temperature, mixing, gas production, and recycle.

One of the principal difficulties experienced has been obtaining sufficient and economical sludge recycle. A sludge degasifier was used in all investigations to separate entrained gases from the anaerobic floc before recycle of the floc to the contact chamber. Although this method proved satisfactory, it would be expensive for a full scale plant.

In later work, Schroepfer and Ziemke (12) have used the findings of Michaelis and Menten (13,14) and Monod (15,16) to develop theoretical equations to explain the anaerobic contact process. Andrews and Pearson (1), Stewart, Pearson, and Hiramoto (17) and Gates et al. (13) have also developed similar approaches for evaluating the anaerobic contact process. All except Gates, however, have only provided the most basic process equation. What is vitally needed, is a detailed evaluation of these theoretical equations in such a manner that they can be applied practically to understanding and controlling the anaerobic contact process.

Gates et al. (13) have conducted an extensive theoretical and experimental investigation of the anaerobic contact process. Monod's (15,16) approach to bacterial growth kinetics was utilized to develop theoretical performance equations for the anaerobic contact process and these equations were evaluated for hypothetical process parameters. Furthermore, the Monod model was used to evaluate data obtained from the operation of a continuous flow, anaerobic-contact reactor. The results were then compared with those obtained by evaluating the



theoretical equations. To further substantiate the findings, the data of Schroepfer et al. (10) and Schroepfer and Ziemke (11,12) were also evaluated and compared with the theoretical.

The use of an upflow type anaerobic sludge separation device provided 95-99 per cent solids separation capabilities at an hydraulic loading rate of 130 gal/ft<sup>2</sup>/day.

Gates et al. (3) have concluded that the Monod model can successfully be applied to an evaluation of process performance for an anaerobic contact process. A method for a determination of the controlling biological growth rate parameters as well as equations for predicting process performance as a function of the controlling variables has been developed. The upflow type clarifier provided satisfactory solids separation and required no operating costs.

However, since the process performance equations were evaluated for hypothetical process parameters, they only provide an indication of process performance and cannot be used to evaluate an actual system. Furthermore, the effluent from an anaerobic contact process is of such quality as to require additional treatment by an aerobic polishing process. There is no literature available to indicate the effect of an anaerobic system effluent on an aerobic system and on the controlling biological parameters. Although Gates et al. indicate that an anaerobic-aerobic type treatment facility may have an economic advantage in the transition region (800 mg/L BOD<sub>5</sub> to 1 per cent organic solids), no evidence of such is presented.

### Thesis Objective

The objective of this thesis was to further investigate the theoretical, economic, and practical feasibility of an anaerobic-aerobic treatment facility.

The work of previous investigators was utilized to develop theoretical process performance equations and graphs for the anaerobic contact process and the activated sludge process. The development was such that the equations can be directly applied to the performance of an anaerobic-aerobic treatment facility.

A method for an economic comparison of the anaerobic-aerobic process with the aerobic process was developed so that the relative economic merit of each process can be evaluated.

A laboratory scale, anaerobic-aerobic reactor was constructed and operated in an attempt to maintain a functioning treatment facility and to substantiate the theoretical process performance equations.

## CHAPTER II

### GENERAL DEVELOPMENT OF REACTION KINETICS AND PROCESS PERFORMANCE EQUATIONS

There are two basic approaches which may be used in the design of a completely mixed activated sludge system. The difference between the two approaches is whether or not the rate of substrate removal from a biological system is a continuous function of organism concentration.

The first approach, presented by McKinney (18), indicates that below a certain food to microorganism ratio, the rate of substrate removal is dependent only on the substrate concentration. He has combined this concept with material balances to develop design equations for activated sludge.

The basic theory for the second approach was presented by Monod (15,16) and applied by Andrews and Pearson (1) and Gates (3,6), while modifications of it have been employed by Eckenfelder and O'Connor (18,19) and Stewart and Ludwig (20). This approach differs from that of McKinney (18) in that substrate removal is assumed to be a function of organism concentration. Since substrate removal is accomplished by conversion of organic matter into microorganism biomass and various products of their metabolic activity, it is reasonable to assume that the rate of substrate removal will be some function of the organism concentration. Since most state and regulatory agencies as well as numerous researchers (1,3,15,20) generally accept that organism

concentration has a continuing influence on substrate removal, the approach of Monod, Gates, Ludwig, etc. will be utilized in this thesis.

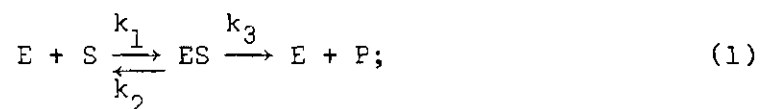
In order to develop predictive equations for the operation of an activated sludge unit, one must first determine the relationship between substrate removal and organism concentration; then using this relationship, evolve a mathematical formulation for the removal of substrate through the activated sludge unit as a function of the appropriate controlling parameters.

By applying the work of Michaelis-Menten in enzyme kinetics (13) and Monod (15) in bacterial growth kinetics and utilizing the concept of the Law of the Minimum, the required constants and parameters for the activated sludge process should be forthcoming.

The Law of the Minimum for a biological population states that development of a population is essentially regulated by the one required substance occurring in minimal quantity relative to the requirement of the population. It may also be applied to chemical reactions. This means that the influence of any controlling parameter on growth will be a maximum at low concentrations and decrease as concentration increases until its effect is negligible. Implicit in this concept is a mathematical relationship whose slope is a maximum at the origin, then decreases as the abscissa value increases, approaching zero at large values of the abscissa.

#### Reaction Kinetics

Michaelis-Menten (13) proposed that enzyme-substrate reactions take place as follows:



where: E = Enzyme.

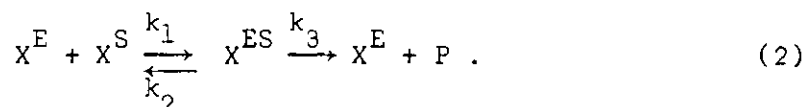
S = Substrate.

ES = Enzyme-Substrate Complex.

P = Reaction Product.

for a steady-state or equilibrium condition.

An enzyme-substrate reaction can be represented by the following relationship where  $X^E$  is the total enzyme concentration in the system,  $X^S$  is the substrate concentration, and  $X^{ES}$  is the enzyme-substrate complex concentration. Assuming a one-to-one interaction between enzyme and substrate, then:



Therefore, the rate of complex formation is

$$\frac{dX^{ES}}{d\theta} = k_1 (X^E - X^{ES})(X^S) \quad (3)$$

and the rate of complex breakdown is

$$\frac{-dX^{ES}}{d\theta} = k_2 X^{ES} + k_3 X^{ES}. \quad (4)$$

When the system is at steady state, the value of  $X^{ES}$  remains constant with time. Thus,

$$\frac{-dX^{ES}}{d\theta} = \frac{dX^{ES}}{d\theta} . \quad (5)$$

Therefore:

$$k_2 X^{ES} + k_3 X^{ES} = k_1 (X^E - X^{ES})(X^S) \quad (6)$$

and

$$\frac{k_2 + k_3}{k_1} : \frac{(X^E - X^{ES})(X^S)}{X^{ES}} = K_M . \quad (7)$$

Equation (7) is the Michaeli-Menten Constant. The rate of conversion of substrate to product is

$$\frac{dX^S}{d\theta} = k_3 X^{ES} = V ; \quad (8)$$

where  $V$  is the velocity of the overall reaction. When the concentration of  $X^S$  is high, the concentration of uncombined enzyme will be zero and the rate of conversion will be a maximum. Thus:

$$X^E = X^{ES} \quad (9)$$

and

$$V_{\max} = k_3 X^E = k_3 X^{ES} . \quad (10)$$

Solving Equation (8) for  $X^{ES}$  and substituting into Equation (7) yields

$$V = k_3 \left[ \frac{X^E X^S}{K_M + X^S} \right]$$

and

$$V = \frac{V_{\max} X^S}{K_M + X^S} . \quad (11)$$

Equation (11) is the Michaelis-Menten equation which states that the velocity at which substrate is converted to product is at any time determined by the maximum velocity at which conversion can take place, the substrate concentration at any time, and the Michaelis-Menten constant. Equation (11) is of the form of a rectangular hyperbola with translated origin and has the same characteristics as required to satisfy the previously described requirements of the Law of the Minimum.

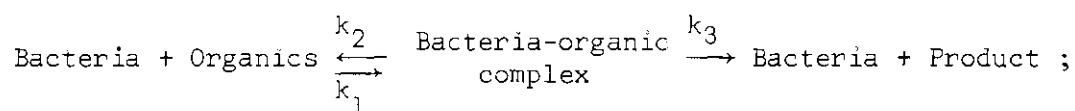
Monod (15) observed that the growth rate constant of his bacterial systems could be expressed as a function of substrate concentration. The form of this expression is

$$k = \frac{k^m X^S}{K + X^S} ; \quad (12)$$

where  $k$  is the biological growth rate constant,  $k^m$  is the maximum growth rate possible by the biological organisms,  $K$  is the concentration of nutrients at which the growth rate is half of the maximum, and  $X^S$  is the concentration of the limiting nutrient in the system.

By assuming that the conversion to end product of the substrate in a bacterial system follows a reaction sequence similar to that of the enzyme kinetics advanced by Michaelis-Menten, the Monod Constant can be shown to be similar in form to that of the Michaelis-Menten constant.

Assume the following reaction occurs:



where:  $X^O$  = organism concentration (total).

$X^N$  = organic concentration (nutrient concentration).

$X^{ON}$  = organism-nutrient complex concentration.

$X^O - X^{ON}$  = organism concentration not involved in complex.

Unless organisms and organics combine in a one-to-one rate,  $X^{ON}$  must be multiplied by an appropriate constant to be used in an actual numerical situation. Thus, it is used symbolically here.

Assuming steady state and applying the procedure followed for the Michaelis-Menten formulation, one obtains the following. At steady state,

$$\frac{dX^O}{d\theta} = 0. \quad (13)$$

The rate of complex formation is

$$\frac{dX^{ON}}{d\theta} = k_1 (X^O - X^{ON}) X^N \quad (14)$$



and the rate of complex breakdown is

$$\frac{-dX^{ON}}{d\theta} = k_3(X^{ON}) + k_2(X^{ON}) . \quad (15)$$

Since the rate of complex formation equals the rate of breakdown.

$$\frac{-dX^{ON}}{d\theta} = \frac{dX^{ON}}{d\theta} . \quad (16)$$

Therefore,

$$k_3X^{ON} + k_2X^{ON} = k_1(X^O - X^{ON})(X^N) \quad (17)$$

and

$$\frac{k_2 + k_3}{k_1} = \frac{(X^O - X^{ON})(X^N)}{X^{ON}} = K . \quad (18)$$

Equation (18) is similar in form to the Michaelis-Menten constant.

The rate of nutrient conversion to "product" is

$$\frac{-dX^N}{d\theta} = k_3X^{ON} = V ; \quad (19)$$

where V is the velocity of conversion, i.e. rate of conversion. If the value of  $X^N$  is sufficiently high such that all the organisms are involved in the complex, then

$$X^O = X^{ON} \quad (20)$$

and

$$\frac{dX^N}{d\theta} = k_3 X^O = V_{\max} ; \quad (21)$$

where  $V_{\max}$  represents the maximum rate of substrate conversion that could be theoretically realized in the system. Substituting Equations (19) and (21) into Equation (18), one obtains:

$$V = \frac{V_{\max} X^N}{K + X^N} ; \quad (22)$$

which is the same in every respect to the expression developed by Michaelis-Menten. It can also be shown that the rate of biological growth can, at any time, be represented by

$$\frac{dX^O}{d\theta} = kX^O . \quad (23)$$

Let  $Y^O = \text{gross yield} = \frac{\text{mg. organism produced}}{\text{mg. nutrient removed}} .$

Then Equation (23) becomes

$$\frac{dX^O}{d\theta} = Y^O \left[ \frac{-dX^N}{d\theta} \right] \quad (24)$$

Since  $V = \frac{-dX^N}{d\theta}$  from Equation (19),

$$\frac{dX^0}{d\theta} = Y^0 V = kX^0 \quad (25)$$

and

$$V = \frac{k}{Y^0} X^0 \quad (26)$$

and

$$V_{\max} = \frac{k_{\max}}{Y^0} X^0. \quad (27)$$

Substituting Equations (26) and (27) into Equation (22) yields

$$k = \frac{k^m X^N}{K + X^N}, \quad (28)$$

which is identical to the Monod constant. However, the following observations seem appropriate:

It would seem from the preceding that the relationships of Michaelis-Menten, Monod, and the Law of the Minimum share certain common concepts due to the fact that they are all expressible by the same mathematical expression. At present, however, it must be strongly urged that the similarity of form represents a necessary but not sufficient condition to say that all three relationships are equivalent. The Michaelis-Menten expression is a classical chemical kinetic equation, however, the hyperbolic relationship between bacterial growth rate and concentration of control parameters should, for the present, be considered as a "process kinetic" relationship, thus even though it is identical to the Michaelis-Menten equation it does not necessarily imply the existence of the same basic principles nor is it to be considered equivalent in terms of relative merit. (24)

In summary, it has been shown that the Monod constant for bacterial systems is similar in form to that advanced by Michaelis-Menten for enzyme kinetics. It seems plausible, therefore, that by evaluating

the appropriate bacterial growth rate constants, relating growth rate to substrate removal, and applying material balances to a bacterial system, the rate of substrate removal and the rate of biomass growth should be predictable. Therefore, the following generalized system will be considered.

#### Process Development for Generalized Treatment System

Let the system be continuous flow in nature and consist of a complete-mix biological reactor equipped with a conventional gravity settler and sludge recycle capabilities. The flow diagram and nomenclature is as follows:

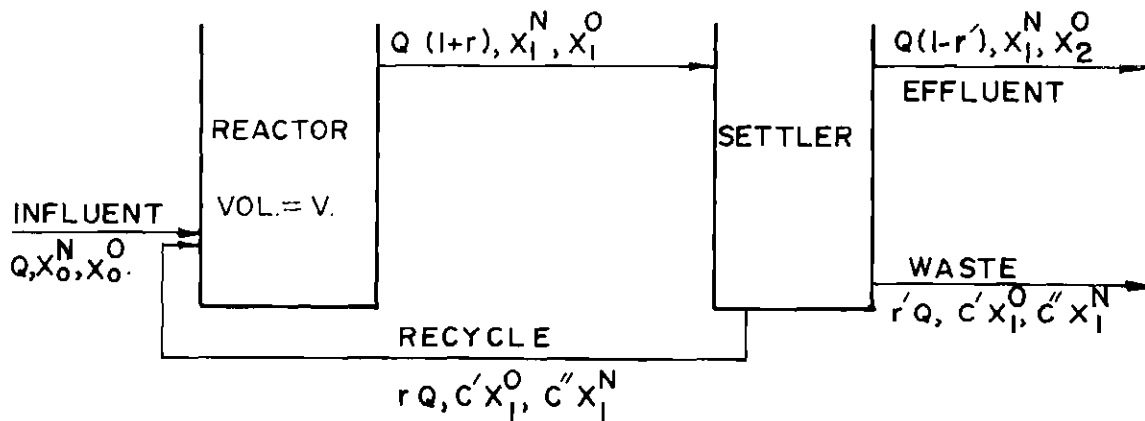


Figure 1. General Treatment System Schematic.

Where:  $Q$  = Reactor influent flow.

$X_0^N$  = Influent nutrient (substrate) concentration.

$X_0^O$  = Influent organism (biomass) concentration.

$V$  = Reactor volume.

$X_1^N$  = Reactor nutrient concentration.

$X_1^0$  = Reactor organism concentration (MLVSS).

$X_2^N$  = Settler effluent nutrient concentration.

$X_2^0$  = Settler effluent organism concentration.

$r$  = Fraction of influent recycled.

$r'$  = Fraction of influent flow to waste.

$C'$  = Degree of concentration of organisms attained in solids separation unit, ( $X^0$  underflow/ $x^0$  settler influent).

$C''$  = Degree of concentration of nutrients attained in the solids separation unit, ( $X^N$  underflow/ $X^N$  settler influent).

If one assumes that bacterial growth always takes place in an exponential fashion with the value of the growth rate constant being dependent on environmental parameters, then the rate of growth will be determined by the value of the growth rate constant for the particular system. Therefore, as previously stated in Equation (23), the rate of bacterial growth will be of the form  $dx^0/d\theta = kX^0$ . Consequently, applying the Monod Model and a material balance for the organisms in the reactor, yields:

$$V \frac{dx_1^0}{d\theta} = QX_o^0 + kX_1^0V + rQC'X_1^0 - (1+r) QX_1^0 - k^D X_1^0V ; \quad (29)$$

where  $k^D$  is the rate of endogenous respiration. Assuming steady state conditions and  $X_o^0 \approx 0$ , then

$$0 = kX_1^0 + \frac{rC'X_1^0}{\theta_T} - \frac{(1+r) X_1^0}{\theta_T} - k^D X_1^0 \quad (30)$$

and

$$k = k^D + \frac{1}{\theta_T} \{ (1 + r(1-C')) \}. \quad (31)$$

Letting  $k = k^m X_1^N / (K + X_1^N)$ , which is the Monod constant, and simplifying expression (31), one obtains:

$$X_1^N = \frac{K \{ k^D \theta_T + [1 + r(1-C')] \}}{k^m \theta_T - \{ k^D \theta_T + [1 + r(1-C')] \}}, \quad (32)$$

which expresses the value of  $X_1^N$  (substrate concentration in the reactor at steady state) when recycle is employed. Note that  $C'$  has a minimum value of 1 at which the settler accomplishes no increase in solids concentration.

Applying the same approach to a material balance for the nutrients in the reactor, one obtains:

$$V \frac{dX_1^N}{d\theta} = QX_o^N - \frac{kX_1^O V}{Y^O} - (1+r) QX_1^N + rQC''X_1^N. \quad (33)$$

Assuming steady state and applying Equation (31) yields:

$$0 = \frac{X_o^N}{\theta_T} - \frac{X_1^O}{Y^O} \left\{ k^D + \frac{1}{\theta_T} [1 + r(1-C')] \right\} - \frac{(1+r)X_1^N}{\theta_T} + \frac{rC''X_1^N}{\theta_T}. \quad (34)$$

Upon simplifying Equation (34), one obtains:

$$X_1^0 = \frac{Y^0 \{X_0^N - X_1^N [1 - r(C''-1)]\}}{k^D \theta_T + [1 + r(1-C')]} , \quad (35)$$

which is the expression for organism concentration ( $X_1^0$ ) at steady state with recycle being employed. It should be noted that in most waste water treatment processes, there will be no increase in nutrient concentration in the settler; therefore,  $C'' = 1.0$ . The variables are  $X_1^N$ ,  $X_1^0$ ,  $\theta_T$ ,  $r$ , and  $C'$ . As is shown,  $\theta_T$ ,  $X_1^N$ ,  $r$ , and  $C'$  are all related. The value of  $C'$  is determined by the settling characteristic of the sludge and can be estimated from the sludge volume index. Thus, selecting any two of the remaining three ( $X_1^N$ ,  $\theta_T$ , and  $r$ ), fixes the system.

A material balance through the settler for the biomass yields:

$$Q(1+r)X_1^0 = rQC'X_1^0 + r'QC'X_1^0 + Q(1-r')X_2^0 . \quad (36)$$

Assuming essentially no biomass in the effluent, i.e.  $X_2^0 \approx 0$ , then

$$(1 + r) = rC' + r'C' \quad (37)$$

and

$$r'C' = 1 + r(1 - C') . \quad (38)$$

Letting  $C^0$  equal the recycle factor, or, the decimal fraction of influent biomass to the settler which leaves the settler in the waste flow, then  $C^0$  can be represented by:

$$C^0 = \frac{X_W^0}{X_1^0} . \quad (39)$$

The average solids loss from the settler can be represented by:

$$X_W^0 = \frac{(1 - r')QX_2^0 + r'QC'X_1^0}{Q} . \quad (40)$$

Again assuming no solids loss in the effluent, then

$$X_W^0 = r'C'X_1^0 \quad (41)$$

and

$$C^0 = \frac{r'C'X_1^0}{X_1^0} = r'C' . \quad (42)$$

From Equation (38)

$$C^0 = 1 + r(1 - C') = r'C' . \quad (43)$$

Therefore, the two basic process equations are derived from Equations (32) and (35) and are:

$$X_1^N = \frac{K(k^D\theta_T + C^0)}{k^m\theta_T - (k^D\theta_T + C^0)} \quad (44)$$

and



$$X_1^O = \frac{Y^O(X_O^N - X_1^N)}{k^D \theta_T + C^O} \quad (45)$$

for reactor nutrient concentration and reactor organism concentration at steady state, respectively. Use of these equations should permit one to predict the performance of a biological system when the controlling parameters are known.

Andrews and Pearson (4) and Gates et al. (5) have shown that the Monod Model is applicable to an anaerobic contact process. Therefore, applying an approach similar to that for the development of equations (44) and (45) one can develop the controlling equations for the anaerobic contact process and develop graphs of process performance as functions of various controlling parameters.

## CHAPTER III

THEORETICAL DEVELOPMENT OF PROCESS PERFORMANCE  
EQUATIONS FOR AN ANAEROBIC-AEROBIC TREATMENT FACILITY

Anaerobic Contact Chamber

Consider an anaerobic contact reactor equipped with sludge wasting capabilities and an upflow clarifier so that there is essentially total recycle of sludge from the settler to the reactor. Such a system can be represented as follows:

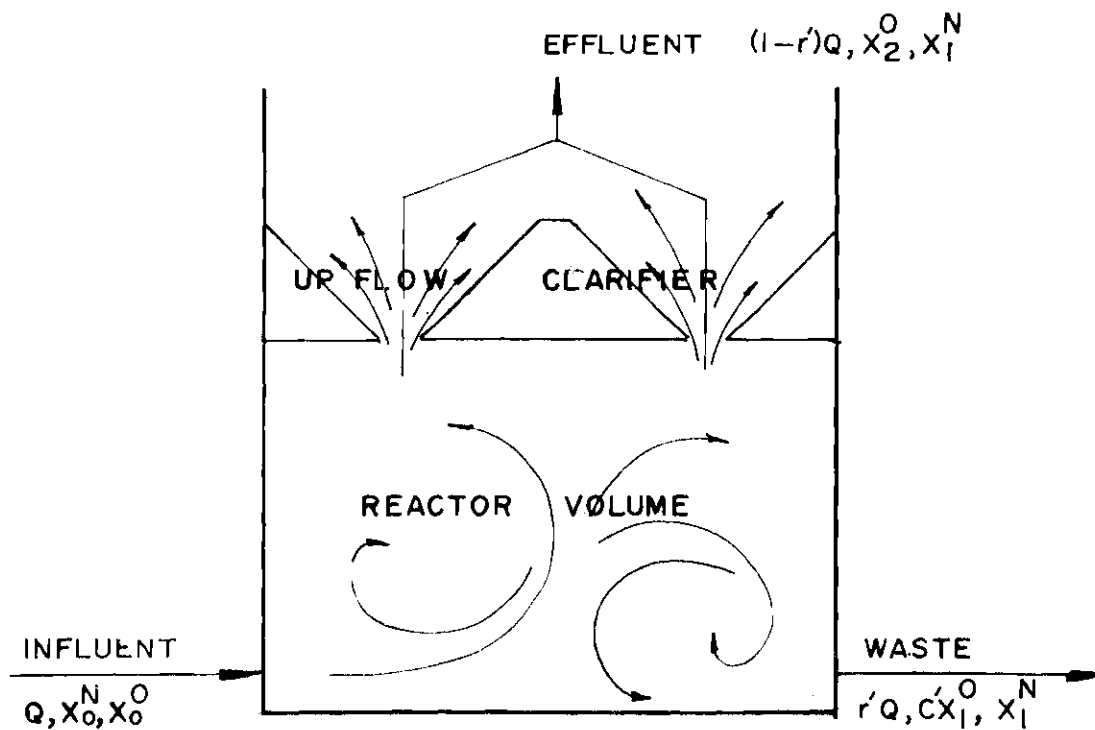


Figure 2. Anaerobic Contact System Schematic.

Applying an organism material balance for the reactor and clarifier yields:

$$V \frac{dX^0}{d\theta} = QX_o^0 + kX_1^0V - k^D X_1^0 V - r'QC'X_1^0 - (1 - r') QX_2^0. \quad (46)$$

By assuming steady state conditions, applying the Monod growth rate constant (Equation (28)) and assuming that the organism concentration in the influent and effluent is approximately zero, one obtains:

$$X_1^N = \frac{K(k^D \theta_T + C^0)}{k^m \theta_T - (k^D \theta_T + C^0)}. \quad (47)$$

Applying a similar procedure for the nutrient material balance yields an equation identical to Equation (45) and is

$$X_1^O = \frac{Y^O(X_o^N - X_1^N)}{k^D \theta_T + C^0}. \quad (48)$$

Therefore, the process performance for the anaerobic contact reactor can be evaluated from Equation (47) and (48) when the appropriate parameters are known. Gates et al. (3) evaluated the bacterial growth rate constants for an anaerobic contact reactor using a synthetic waste water composed of skimmed dried milk in the temperature range of 20-25°C. The values obtained were:

$$k^m = 0.14 \text{ day}^{-1}.$$

$$k^D = 0.07 \text{ day}^{-1}.$$

$$K = 24.3 \text{ mg/L.}$$

$$Y^O = 0.37 \text{ mg VSS/mg COD.}$$

By applying the above constants to Equation (47), a family of curves can be developed for reactor nutrient concentration as a function of the recycle fraction. It has already been noted that  $C^O$  has a maximum value of one and a minimum value of zero at which there is no sludge recycle, respectively. A plot of such a family of curves for  $C^O$  values of 0.02 to 0.10 appears in Figure 3.

It should be remembered, that since the reactor is completely mixed, the value of  $X_1^N$  is also the value of effluent nutrient concentration. On examination of Equation (47) and Figure 3, it is evident that  $X_1^N$  is independent of  $X_O^N$  and is a function only of recycle, bacterial growth parameters, and detention time. As the detention time decreases,  $X_1^N$  increases in an exponential fashion. However, the limiting value of  $X_1^N$  as  $\theta_T$  approaches zero is  $X_O^N$  at which point there is no nutrient reduction and, consequently, no bacterial growth, which would necessitate no biomass. At this critical detention time, total washout of the biomass would occur. By assuming a value for  $X_O^N$  and substituting into Equation (49), the critical detention time can be determined.

$$\theta_{T_c} = \frac{C^O(K + X_1^N)}{k^m X_1^N - k^D(K + X_1^N)} \quad (49)$$

A value of 1000 mg/L COD was assumed for the influent nutrient concentration ( $X_O^N$ ) to develop the curves of Figure 3. Therefore, the critical detention time for washout would be that detention time in Figure 3 where  $X_1^N = 1000 = X_O^N$  for the particular recycle fraction.

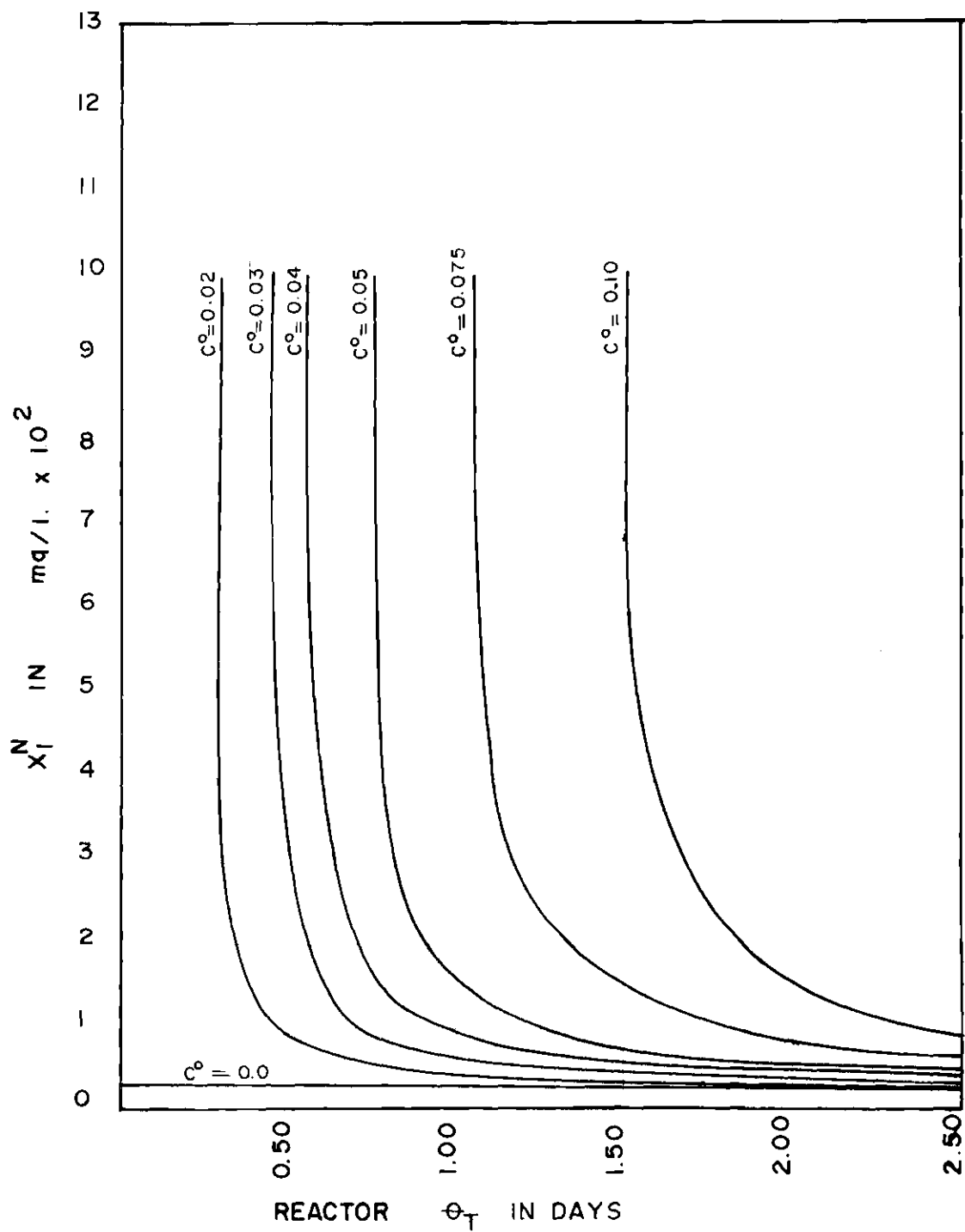


Figure 3. Reactor Nutrient Concentration Versus Theoretical Detention Time for the Anaerobic Contact Process for Various Recycle Fractions.

It can also be seen from Figure 3 that as  $\theta_T$  approaches infinity, the effect of recycle becomes negligible and  $X_1^N$  approaches a constant value. This limiting value may be determined by letting  $C^0$  equal zero, i.e. complete recycle of the anaerobic sludge. Under such conditions, from Equation (31) and (43),  $k = k^D$  and

$$X_1^N = \frac{K k^D}{k^m - k^D} . \quad (50)$$

Evaluating Equation (50) yields the minimum possible effluent nutrient concentration obtainable by the system. Using the constants developed by Gates et al. (3), the minimum effluent concentration is 24.3 mg/L COD and is represented in Figure 3 at a  $C^0$  value of 0.0. Sample computations for the development of the curves of Figure 3 are presented in Appendix B, Table B-1.

Thus, by evaluating the recycle fraction and the theoretical detention time for the particular system, one should be able to predict the effluent nutrient concentration and process performance. Knowing  $X_1^N$  and  $X_0^N$ , then an evaluation of the required organism concentration can be made.

The required organism concentration for steady state can be evaluated for Equation (48), which is

$$X_1^O = \frac{Y^O (X_0^N - X_1^N)}{k^D \theta_T + C^0} .$$

Consequently, the required biomass concentration is a function of the bacterial growth parameters, the theoretical detention time, the recycle

fraction, and the substrate concentration reduction through the unit. Should the biomass concentration be initially greater or less than that required, then there will be a net biological death or growth, respectively, until the nutrient available is just sufficient to maintain the required population, and steady state conditions are realized.

The reactor substrate concentration can be evaluated from Figure 3. Therefore, by evaluating  $X_O^N$ ,  $C^O$ , and  $\theta_T$  for the actual plant, one can determine the required organism concentration from Equation (48). A family of curves of  $X_1^O$  as a function of  $\theta_T$  for various recycle fractions is presented in Figure 4, and sample calculations in Appendix B, Table B-2.

As can be seen from Figure 4, the organism concentration becomes less dependent on recycle as detention time increases. The detention time at which the organisms concentration is equal to zero represents the theoretical washout time. The ability of the system to prevent washout increases as the value of  $C^O$  decreases.

The overall performance of a reactor and the reactor response to any variable is readily determined by consideration of the rate of substrate conversion per unit volume, or:

$$\frac{(X_O^N - X_1^N)}{\theta_T} = \frac{1}{\theta_T} \left[ X_O^N - \frac{K(k^D \theta_T + C^O)}{k^m \theta_T - (k^D \theta_T + C^O)} \right] . \quad (51)$$

$X_1^N$  can readily be determined from Figure 3 for any value of  $C^O$  and  $\theta_T$ . Figure 5 represents a family of curves for the rate of substrate

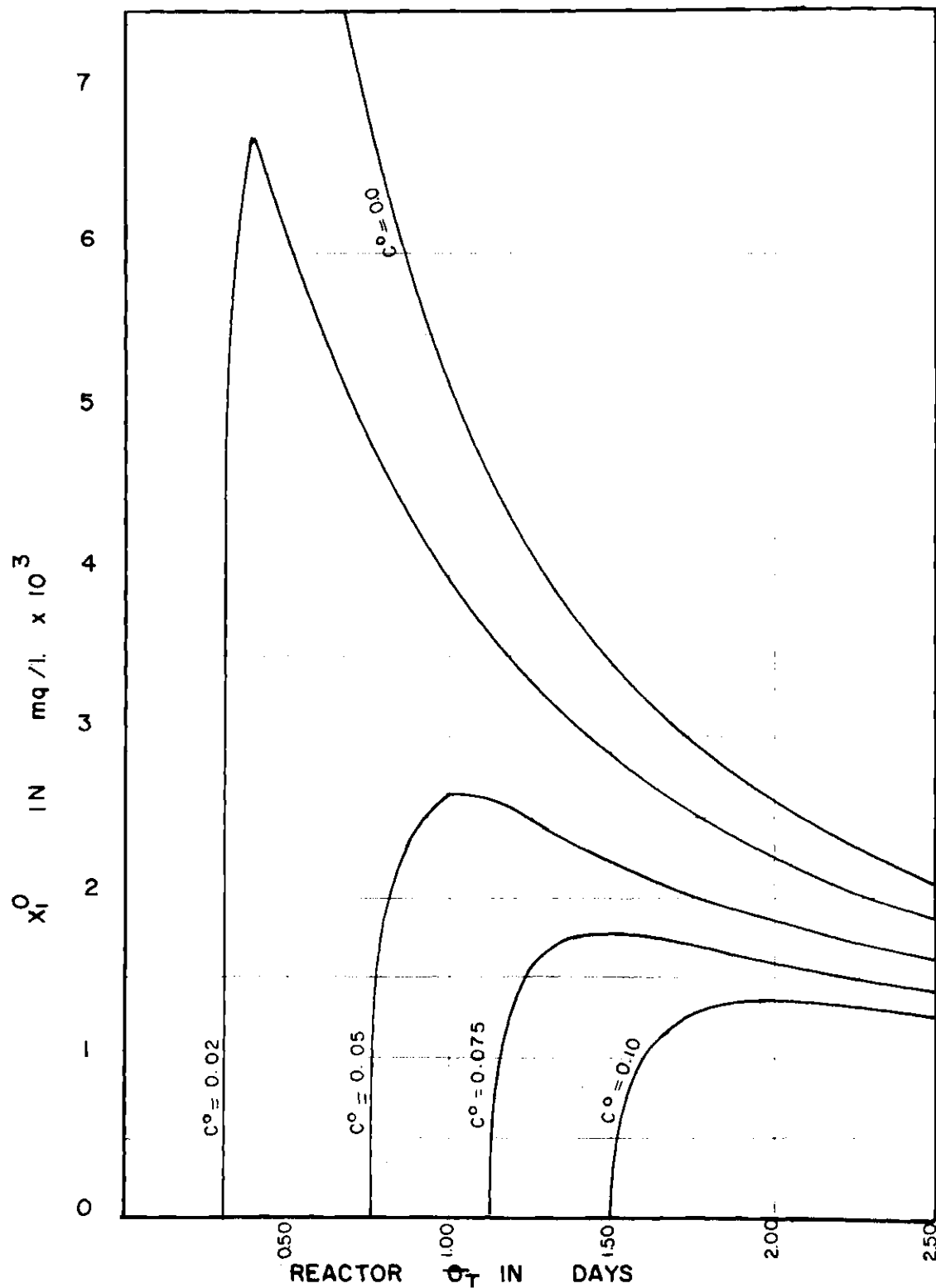


Figure 4. Required Organism Concentration Versus Theoretical Detention Time for the Anaerobic Contact Process for Various Recycle Fractions.



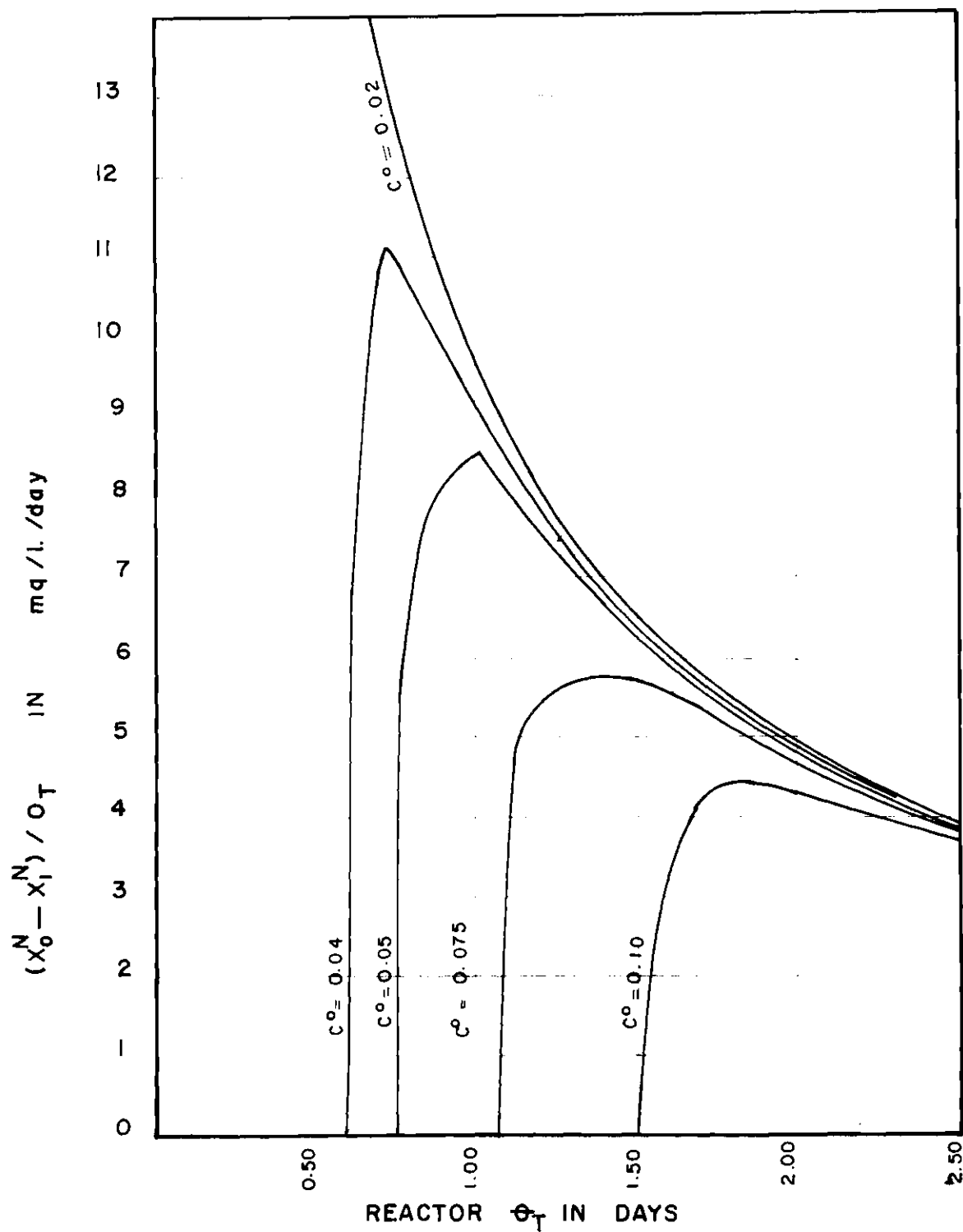


Figure 5. Rate of Substrate Reduction Per Unit Volume as a Function of Detention Time for the Anaerobic Contact Process for Various Values of Recycle Fractions.

removal as a function of  $\theta_T$  for various recycle fractions. There is a definite increasing influence of recycle with decreasing detention time. This is evidence of the compensating effect of recycle for low growth rates.

Obviously, it would be ideal to operate a facility at or near the detention time for maximum substrate removal. For example, a detention time of 1.0 days at a  $C^0$  value of 0.05 will yield the maximum rate of substrate removal. However, it should be pointed out most emphatically that a slight decrease in theoretical detention time from this point, will result in washout and system failure. Therefore, whenever operating near the critical detention time, process control must be rigidly maintained. Also a system operated at this point may not provide the effluent quality required. Sample calculations for Figure 5 are presented in Appendix B, Table B-3.

In summary, by evaluating the controlling parameters of the system, i.e.  $X_O^N$ ,  $C^0$ ,  $\theta_T$ , the process performance can be predicted from Figures 3, 4, and 5; assuming of course, that the biological growth rate constants are the same as those presented. Should this not be the case, then one would necessarily be required to evaluate the growth rate constants for the particular system and the particular substrate and then develop new curves utilizing the new constants. A method for the evaluation of these constants is presented by Gates (21) and Gates et al. (3).

Because of low detention times or highly concentrated waste water, it may be desirable to give the effluent from the anaerobic contact process additional treatment before discharge to the receiving

waters. The passage of the anaerobic effluent through a completely mixed, activated sludge unit offers one possibility.

In such a system, a substantial reduction in BOD may be accomplished at low detention times. Aeration of the anaerobic effluent before discharge will also eliminate much of the initial COD by oxidation of inorganic compounds such as  $\text{NH}_3$ ,  $\text{NH}_4^+$ ,  $\text{NO}_2^-$ , and  $\text{H}_2\text{S}$  as well as considerable odor control by oxidation of the hydrogen sulfide.

The Monod constant was developed for the aerobic system and has been principally applied to aerobic systems (15,19,20,21). Therefore, the same approach as has been applied to the anaerobic contact process can be applied to the completely mixed activated sludge process.

#### Complete Mix Activated Sludge Unit

Rather than changing the nomenclature at this point, the same nomenclature as used in the development of the anaerobic contact design will be applied to the activated sludge design. Let it be emphasized, however, that even though the nomenclature is the same, the values of the controlling parameters are not the same, and the influent to the aerobic unit, for the system envisioned, would be the effluent from the anaerobic unit.

Consider an activated sludge unit similar to that of the anaerobic unit. The reactor is continuous flow, completely mixed, has an up-flow clarifier, sludge recycle, and sludge wasting.

Such a system may be represented by:

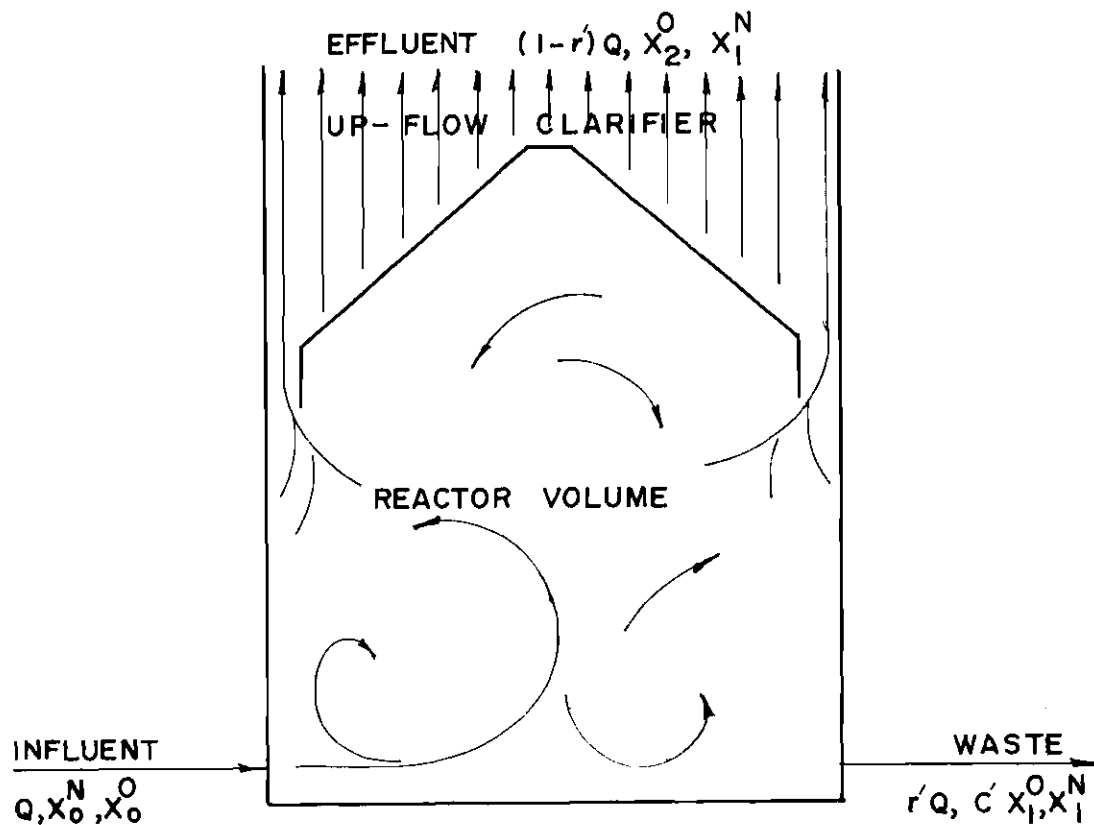


Figure 6. Activated Sludge System Schematic.

Applying a material balance for the nutrient concentration in the reactor will yield:

$$V \frac{dx_1^N}{d\theta} = QX_0^N - r'QX_1^N - (1 - r')QX_1^N - V k' X_1^N X_1^O. \quad (52)$$

Note that in this equation, the parameter  $k'$  instead of  $k$  (as was previously used in the anaerobic design) is utilized. The reason for this is that there has been no evaluation in the literature for Monod's constant, in the form as previously presented, for an activated sludge

system. However, Stewart and Ludwig (20) have evaluated the constant  $k$  for an activated sludge system as  $0.02 \text{ Day}^{-1}(\text{Mg MLVSS/L})^{-1}$ , and it has been utilized for design by Gates (21).

Applying a material balance for the substrate through the reactor and clarifier, assuming steady state conditions ( $dx^N/d\theta = 0$ ), and using  $Q/V = 1/\theta_T$ , then:

$$0 = \frac{x_o^N}{\theta_T} - \frac{r'x_1^N}{\theta_T} - \frac{(1 - r')x_1^N}{\theta_T} - k'x_1^Nx_1^O. \quad (53)$$

Simplifying Equation (53) yields:

$$x_1^O = \frac{1}{k'\theta_T} \left[ \frac{x_o^N}{x_1^N} - 1 \right]. \quad (54)$$

Applying an organism material balance through the reactor and clarifier yields:

$$V \frac{dx_1^O}{d\theta} = Qx_o^O - r'QC'x_1^O - (1 - r')Qx_2^O + Y^Ok'x_1^Ox_1^NV - k^Dx_1^OV.$$

Assuming that steady state conditions exist, i.e.  $dx^O/d\theta = 0$ , and assuming that the organism concentration in the influent and effluent are essentially zero, then

$$0 = \frac{-r'C'x_1^O}{\theta_T} + Y^Ok'x_1^Ox_1^N - k^Dx_1^O. \quad (55)$$

Solving Equation (55) for  $X_1^N$  and using Equation (43), then

$$X_1^N = \frac{k^D \theta_T + C^O}{Y^O k' \theta_T} \quad (56)$$

Equation (56) and (54) are the equations for the nutrient concentration in the reactor and effluent and the organism concentration in the reactor at steady state, respectively. By evaluation of the necessary parameters in the two equations, the aerobic process performance can be characterized.

As has been previously indicated, the bacterial growth rate constants have been determined by Stewart and Ludwig (20) for an MAS system. These constants, which are utilized in this paper for the theoretical evaluation of the activated sludge system, have the following values:

$$k' = 0.02 \text{ Day}^{-1} (\text{mg MLVSS/L})^{-1}.$$

$$k^D = 0.05 \text{ Day}^{-1}.$$

$$Y^O = 0.60 \text{ mg MLVSS/mg BOD}_5 \text{ Removed}.$$

It can be determined from Equation (56) that the reactor substrate concentration is a function of the bacterial growth parameters, recycle fraction, and theoretical detention time, but is independent of influent nutrient concentration and organism concentration; providing of course that the nutrient is the limiting factor and not the organism concentration. Applying the above constants to Equation (56) yields

$$X_1^N = 4.167 + 83.333 \frac{C^0}{\theta_T} . \quad (57)$$

A family of curves can be developed from Equation (57) by plotting  $X_1^N$  as a function of  $\theta_T$  for different values of  $C^0$ . Such a set of curves is developed in Figure 7 for  $C^0$  values of 0.01 to 0.10. Sample computations for this development are presented in Appendix C, Table C-1.

It is evident from Figure 7 that the value of recycle diminishes as the detention time increases, but recycle becomes more and more important as  $\theta_T$  approaches zero. Obviously, it is desirable to operate at the minimum detention time commensurate with the desired degree of treatment. However, as the detention time decreases, the  $X_1^N$  increases in an exponential fashion. The limiting value of  $\theta_T$  is that value where where the reactor nutrient concentration equals the influent nutrient concentration. Under such a situation, there could be no biological growth since there is no substrate removal. This would necessitate no biomass and washout would occur since the biomass would be washed from the system faster than it could be replaced by growth of additional organisms. By letting  $X_1^N$  equal  $X_o^N$  in Equation (57), the critical detention time may be determined by

$$\theta_{Tc} = 4.167 + 83.333 \frac{C^0}{X_o^N} .$$

An influent substrate concentration of 1000 mg/L was used in developing the curves for Figure 7. The critical detention time is found for the particular  $C^0$  value as that time where  $X_1^N = 1000$  mg/L.

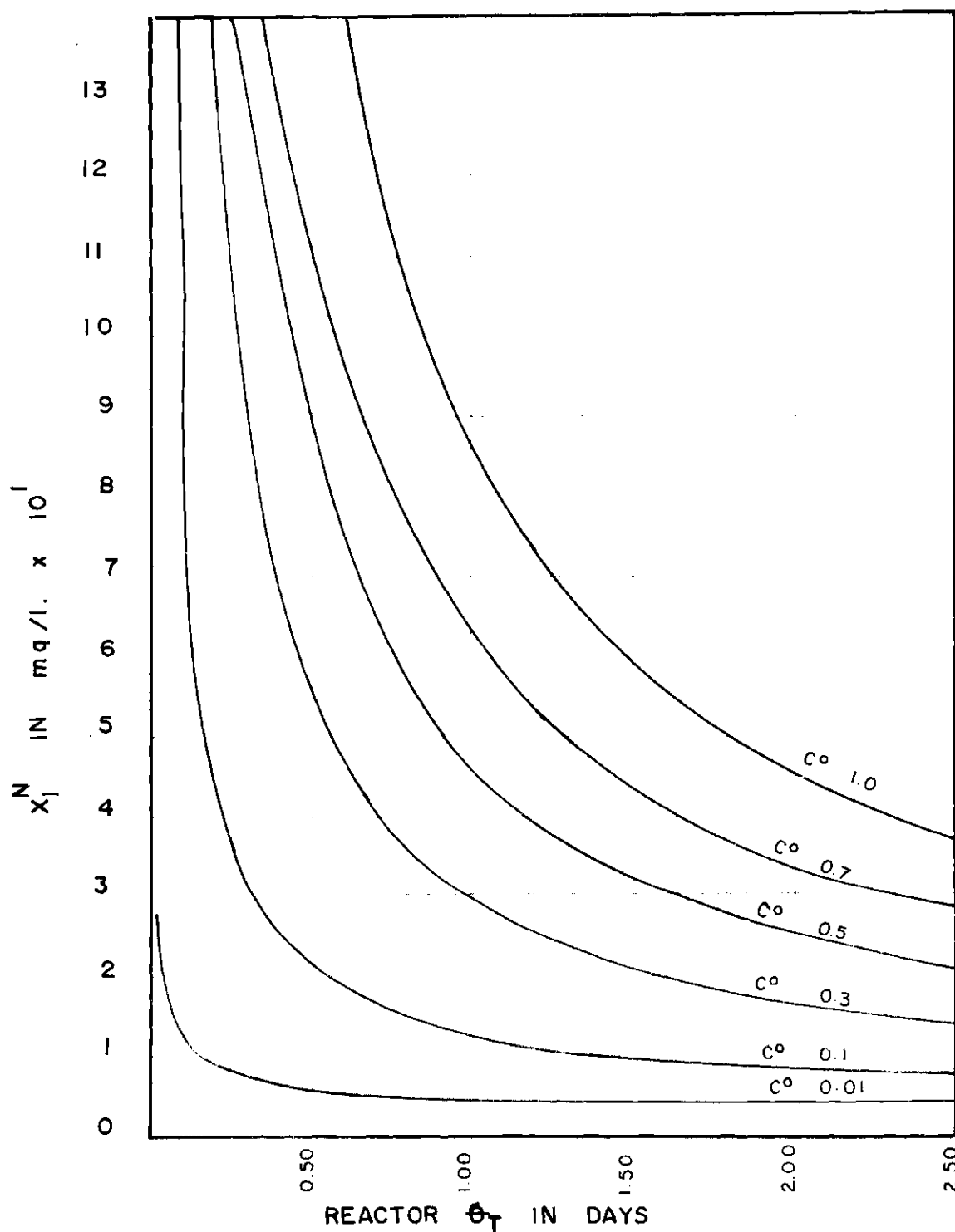


Figure 7. Reactor Substrate Concentration as a Function of Theoretical Detention Time for the Activated Sludge Process for Various Recycle Fractions.



Figure 7 also indicates that a limiting value is approached as  $C^0$  approaches zero. When  $C^0 = 0$ , there is complete recycle and the system stabilizes with growth equalling death, or:

$$Y^O_k X_1^N = k^D ,$$

and from Equation (57) with  $C^0 = 0$ ,

$$X_1^N = 4.167 \text{ mg/L} .$$

Equation (57) indicates that the value of substrate concentration is a function only of the ratio of recycle to detention time and is independent of the discrete values of either variable. Therefore, for a given ratio, the value of  $X_1^N$  will be constant. A series of curves showing this relationship is presented in Figure 8. For a given value of  $X_1^N$ , any combination of  $C^0$  and  $\theta_T$  can be used to obtain the same result.

Therefore, by evaluating  $C^0$ , and  $\theta_T$  for the actual installation, a determination of  $X_1^N$  can be evaluated from Figures 7 and 8 or Equation (57). Knowing  $X_1^N$  and using Equation (54), the required organism concentration for steady state conditions can be evaluated.

Note that  $X_1^O$  is dependent on  $C^0$  only as it affects the substrate concentration. Applying the constants to Equation (54) yields

$$X_1^O = \frac{50}{\theta_T} \left[ \frac{X_1^N}{X_1^N} - 1 \right] , \quad (58)$$

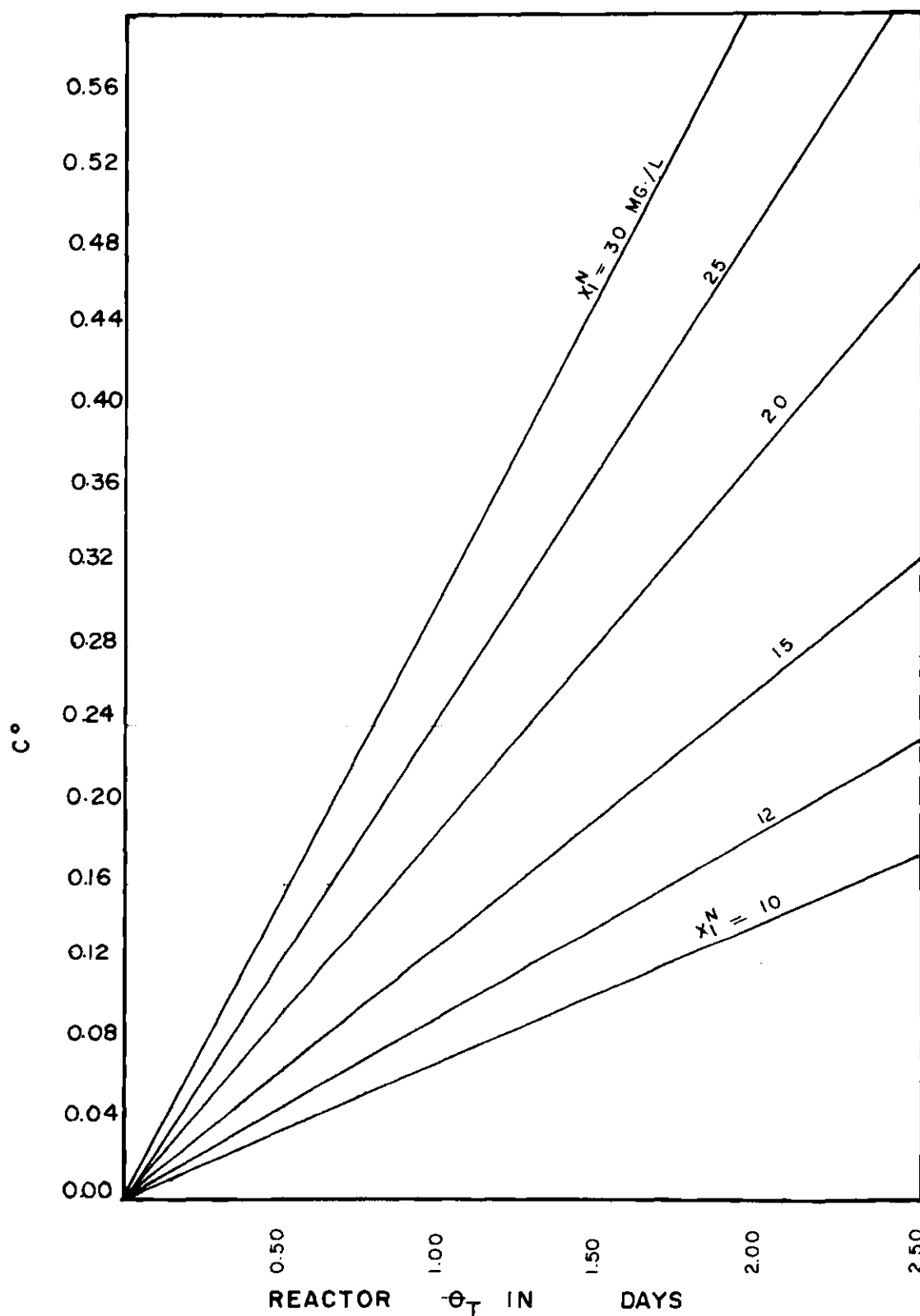


Figure 8. Relationship Between  $C^\circ$  and  $\theta_T$  for the Activated Sludge Process for Various Reactor Substrate Concentrations.

from which a series of curves can be developed to predict the required organism concentration if the influent concentration is set. Since it is desired, in this case, to have the influent concentration to the aerobic section dependent on the effluent concentration from the anaerobic contact process as previously discussed, the graphs for this relationship will not be presented. One should evaluate Equation (58) for the particular situation. It will be stated, however, that the graphs for this analysis would be identical in shape and characteristics, but not in values, to those developed in Figure 4.

The two boundary conditions for the organism concentration are when  $X_1^N$  equals  $X_0^N$  and when  $C^O$  equals zero. For the former, evaluation of Equation (54) shows that the organism concentration is zero, which agrees with the determination made in the evaluation of the limit for reactor nutrient concentration. For the latter condition,

$$X_1^O = \frac{50}{\theta_T} \left[ \frac{X_0^N}{4.167} - 1 \right] .$$

Since  $X_1^O$  is dependent on the ratio of the influent nutrient concentration to the effluent nutrient concentration, which is the reciprocal of the per cent of nutrient remaining, a series of curves can be developed showing the required organism concentration as a function of detention time for any given per cent COD reduction through the aerobic system. Such a series of curves is presented in Figure 9, with sample calculation appearing in Appendix C, Table C-2. Note again that any combination of  $X_1^O$  and  $1/\theta_T$  may be used to obtain a specified degree of treatment.

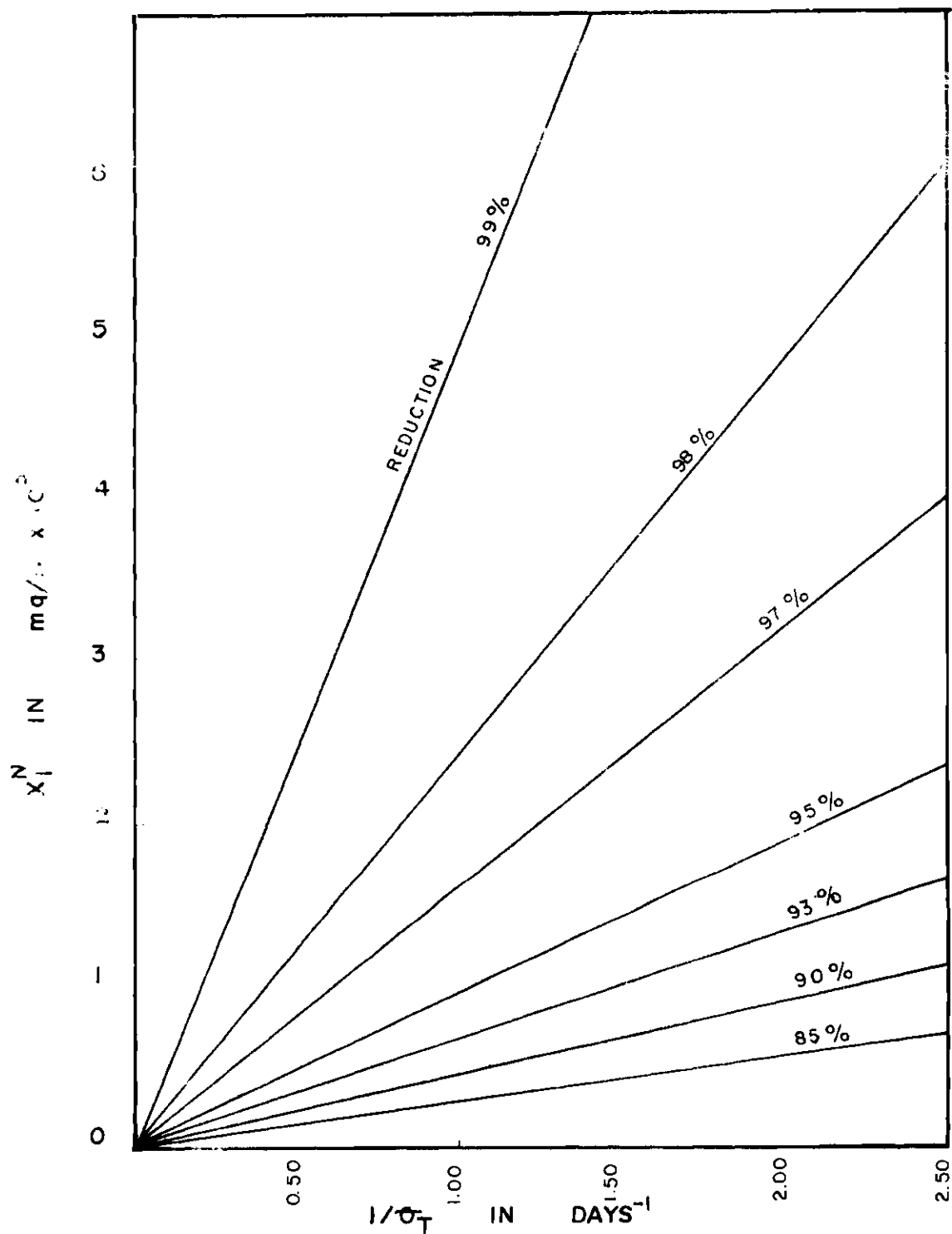


Figure 9. Organism Concentration Versus Detention Time for Various Percent Reductions of Organic Load in an Activated Sludge Unit.

The self-regulating action of a completely mixed biological system is demonstrated in Figures 3, 4, 7 and 9 and by Equations (47), (48), (54), and (56). If  $\theta_T$  remains constant but  $X_O^N$  varies, the effluent quality remains unchanged. If, compared to design conditions,  $\theta_T$  increases and/or  $X_O^N$  increases, an expected condition in waste water treatment, the value of  $X_1^N$  will remain constant or decrease depending on whether  $C^O$  is allowed to increase in proportion to  $\theta_T$  or is held constant.  $X_1^O$  will change in accordance with the changes of  $X_O^N$  and  $\theta_T$ . Thus for flows less than design and BOD concentration greater than design conditions, the performance of the process will equal or exceed design requirements in terms of either per cent removal, effluent quality, or both.

Unfortunately, the system is not self regulating per se when flow rate increases ( $\theta_T$  decreases). This is also a situation commonly encountered in waste water treatment plants. In order to maintain the desired effluent quality for such a situation, the  $C^O$  value must be decreased. Process operation is, therefore, required under such conditions. However, Gates et al. (3) have shown that clarifiers generally used with such reactors demonstrate an ability to maintain  $C^O$  constant, within limits, irrespective of flow rate. Thus, establishing the required  $C^O$  value for maximum conditions will usually result in more than sufficient recycle for all other conditions.

## CHAPTER IV

### COMPARISON OF AN ANAEROBIC-AEROBIC SYSTEM WITH AN AEROBIC SYSTEM ON AN ECONOMIC BASE

#### Introduction

In Chapters I, II, and III a discussion of the anaerobic process and the aerobic process was presented and theoretical equations and graphs were developed to predict the performance of such units under various conditions. However, in any actual application of waste water treatment, the ultimate objective is to obtain the desired degree of treatment for the least cost.

The cost of waste water treatment can be divided into two parts. The first part is the initial cost of the installation which must be amortized over a reasonable period of time. Second is the operating cost which should remain relatively constant from year to year, provided that the waste water characteristics also remain relatively constant.

An inspection of Figures 3 and 7 will show that for equivalent recycle rates and effluent quality, the aerobic process required the least detention time, which is an indication that the initial cost will also be smaller. However, oxygen must be supplied to the aerobic process, which is not the case with the anaerobic process. Thus, the operating cost of the aerobic process will be greater than that of the anaerobic process. Another fact which favors the aerobic process is that the anaerobic process may not be capable of providing the desired

degree of treatment. Consequently, the anaerobic effluent may have to be sent through an aerobic polishing process. This would tend to increase the cost of the anaerobic treatment process.

There is but one method by which to determine the economic advantage to either type of treatment, and that is to independently evaluate each system on an economic base. Obviously, there are an infinite number of variables which can be considered in such an evaluation, but by placing certain restrictions and making a number of assumptions, an economic evaluation for the two systems should be forthcoming.

#### Economic Evaluation

The anaerobic waste treatment process is different in many respects from that of the aerobic process. However, from an economic standpoint, the two systems are quite similar.

Since both the anaerobic and aerobic treatment facilities would require similar pretreatment equipment, chlorination equipment, piping, laboratories, etc.; this evaluation will be limited to those phases of the treatment process for which a distinct economic advantage may be exhibited by one type of treatment. These phases will be those which are most intimately affected by the basic nature of the anaerobic and aerobic processes.

The rate at which stabilization occurs, the yield in biomass, and the operating requirements are the major differences between the anaerobic process and the aerobic process which would most greatly affect the economic evaluation. Consequently, this evaluation can be

limited to three phases of the treatment facility: (1) the size of the reactor chamber which is a function of theoretical detention time, (2) the sludge handling facilities (clarifier, scrappers, returns) which are functionally related to sludge yield, and (3) operating requirements.

In order to evaluate economically these phases of the treatment process, one must make a number of assumptions.

First, since the lower limit of influent  $BOD_5$  for successful operation of the anaerobic-aerobic system is about 1500 mg/L and since the effluent should be of equivalent or better quality than that presently discharged from municipal waste water treatment facilities, the influent organic load will be set at 1500 mg/L  $BOD_5$  and the effluent quality set at 20 mg/L  $BOD_5$ .

Therefore, the following criteria will be used in this evaluation:

- a. Flow = 1MGD = 133,700 cu.ft./day = 5771 cu.ft./hr.
- b. Influent = 1500 mg/L  $BOD_5$ .
- c. Effluent = 20 mg/L  $BOD_5$ .
- d. The constructed cost of reactor and clarifier vessels is \$100.00 per cubic yard of concrete in walls and bottoms.
- e. Liquid depth is 10 feet.
- f. The horizontal cross sections of the reactors are square.
- g. Clarifiers are circular.
- h. All walls and bottoms are 12 inches thick.
- i. The combination of motors and drive mechanisms can be purchased for \$250.00/Hp.



- j. Electricity costs \$0.025/KWH.
- k. Clarifier sludge collectors can be purchased for \$100.00 per foot of diameter exclusive of motor and drive mechanism.
- l. The standard surface loading of clarifiers is 800 gal/sq ft/day for a MLVSS concentration of 2500 mg/L and the surface area must be increased in direct proportion to the ratio of the MLVSS concentration to 2500 mg/L.
- m. The life of the facility is 10 years and the interest rate is 5 per cent.
- n. The annual cost of capital recovery (ACCR) is:

$$ACCR = \left[ \frac{P}{10} \right] + P \left[ \frac{.05}{2} \right] \left[ \frac{10 + 1}{10} \right] = P(0.1275)$$

P = Initial Cost.

- o. Mixing requires 1 Hp-Hr/1000 cu.ft.
- p. Use mechanical mixing which transfers 2.5# O<sub>2</sub>/Hp-Hr.
- q. 2.4 Hp. to drive sludge collectors in clarifiers.

#### Reactor Structure and Mixing Costs

Since both the anaerobic reactor and the aerobic reactor (exclusive of aeration and sludge handling facilities) are subject to the same mixing requirements per unit volume, the basic costs for either type of treatment facility will consist of the cost of initial construction plus the cost of the mixing requirements. The initial construction cost represents the cost of the reactor structure itself (concrete requirements) which must be amortized over the ten-year life of the facility. The volume of the reactor is dependent on the

theoretical detention time. Thus, the initial costs will also be dependent on the detention time.

The mixing requirements have been established as one Hp-Hr/1000 cu.ft. This represents both an initial cost and an operation cost. The initial cost of \$250.00 per horsepower must be amortized over the ten-year life of the facility and represents an annual cost of \$31.87/Hp-Hr. The operating cost is an annual power cost of \$0.01867/Hp-Hr. of operation. Assuming a one horsepower motor running continuously for one year results in an annual operating cost of \$163.50/yr/hp. Therefore, the total annual cost for mixing is \$195.37/1000 ft<sup>3</sup>/yr.

The annual cost per cubic foot of reactor volume for mixing and volume requirements, for both the anaerobic and the aerobic reactors, can be plotted as a function of theoretical detention time. Figure 10 is such a plot. Note that the mixing cost per unit volume as a function of  $\theta_T$  is a constant. The volumetric costs at a low detention time increase rapidly since the ratio of volume to surface area of the reactor is decreasing.

#### Aerobic Treatment Cost

The quantity and rate at which oxygen must be supplied to an activated sludge system are dependent upon the rate at which the biodegradable matter in the waste water is removed, the quantity that is removed, and the rate and quantity of endogenous respiration. Assuming that 0.6 mg. of oxygen are utilized per mg. of BOD<sub>5</sub> removed, the rate of oxygen consumption can be represented by the following equations:

$$\text{Oxygen Consumption} = \text{Endogenous Respiration} + \text{Respiration} .$$

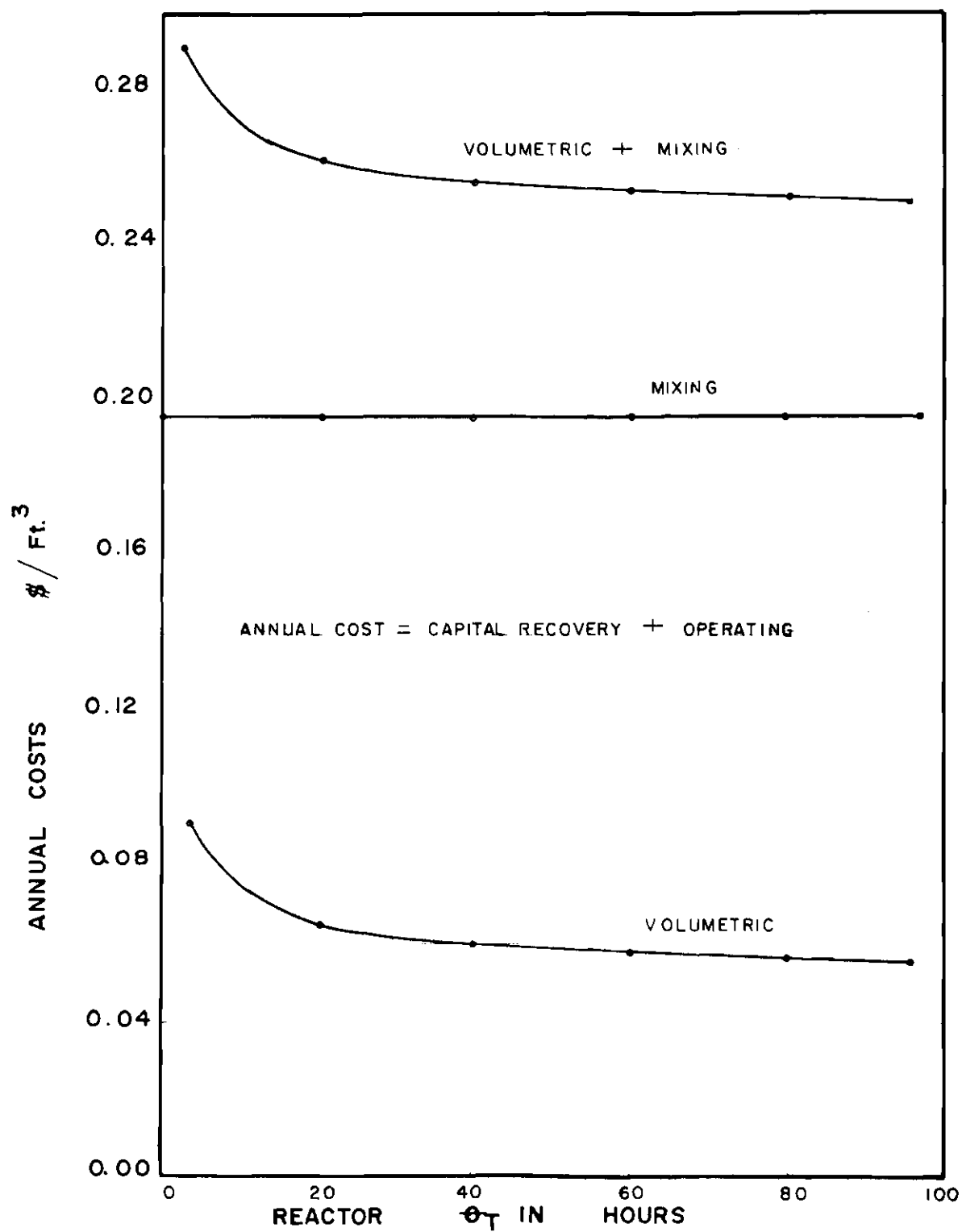


Figure 10. Annual Mixing and Volumetric Costs as a Function of Detention Time.

$$\frac{\text{mg. O}_2/\text{L}}{\text{Day}} = \left[ \frac{X_1^O k^D}{Y_O} + \frac{X_0^N - X_1^N}{\theta_T} \right] \left[ 0.6 \frac{\text{mg. O}_2 \text{ utilized}}{\text{mg. BOD}_5 \text{ removed}} \right] \quad (59)$$

The nomenclature and constants in the above equation are the same as those presented in Chapter III. By substituting Equation (54) for  $X_1^O$ , and since the influent substrate concentration ( $X_0^N$ ) and the effluent substrate concentration ( $X_1^N$ ) are usually set for any treatment facility, the only remaining variable is  $\theta_T$ .

For this analysis, consideration will be given to three cases for the aerobic system. For Case I, aerobic treatment will be the only treatment given to the assumed waste water load. Consequently,  $X_0^N$  will be 1500 mg/L BOD<sub>5</sub> and  $X_1^N$  will be 20 mg/L BOD<sub>5</sub>. For Case II,  $X_0^N$  is 1000 mg/L BOD<sub>5</sub> and  $X_1^N$  is 20 mg/L BOD<sub>5</sub>. For Case III,  $X_0^N$  is 200 mg/L BOD<sub>5</sub> and  $X_1^N$  is 20 mg/L BOD<sub>5</sub>. This last situation is representative of a system where the anaerobic process might be used to stabilize the major waste load and the effluent then passed through an aerobic polishing process to reduce the BOD down to the 20 mg/L maximum, i.e. an anaerobic-aerobic process.

Since all variables in Equation (59), except  $\theta_T$  have now been defined, the rate of oxygen consumption for aerobic stabilization of the waste load can now be determined for the three cases and, consequently, the cost of the stabilization process.

Evaluation of Equation (59) for the three cases results in the following rates of oxygen consumption:

$$\text{Case I} \quad \text{mg. O}_2/\text{L/Day} = \frac{1073}{\theta_T} = 0.06705 \text{ \#O}_2/\text{Ft}^3/\theta_T . \quad (60)$$

$$\text{Case II} \quad \text{mg. O}_2/\text{L/Day} = \frac{711}{\theta_T} = 0.0445 \text{ \#O}_2/\text{Ft}^3/\theta_T . \quad (61)$$

$$\text{Case III} \quad \text{mg. O}_2/\text{L/Day} = \frac{130.5}{\theta_T} = 0.00815 \text{ \#O}_2/\text{Ft}^3/\theta_T . \quad (62)$$

Since the oxygen is supplied by mechanical equipment, the initial cost of the aeration equipment plus the operating cost of the equipment must be considered in the cost of the treatment facility. Oxygen can be supplied at the rate of 2.5 #O<sub>2</sub>/Hp-Hr. and at an operating cost of \$0.01867/Hp-Hr. Since a 1/60 horsepower motor operating continuously will supply 1 #O<sub>2</sub>/Day, the total annual cost of supplying 1 #O<sub>2</sub>/Day will be comprised of \$0.53/hr., ACCR, plus \$2.72/yr., operation cost.

For each of the three cases above, the pounds of oxygen required per cubic foot of reactor volume per day for various detention times can be calculated from Equations (60), (61), and (62) by substitution of the appropriate value of  $\theta_T$ . Once the quantity of oxygen required has been determined, then the total annual operating cost of supplying that oxygen can be calculated at the rate of \$3.25/yr/(#O<sub>2</sub>/Day).

Since the initial cost of the reactor structure is a function only of the detention time, the initial cost for the reaction chamber in each of the three cases would be the same for a given detention time. Consequently, since the volumetric cost (initial reaction chamber cost) has already been determined in Figure 10, one need only

add the initial cost from Figure 10 to the operating cost of supplying oxygen for a particular detention time to obtain the total annual cost of constructing and operating an aerobic reaction chamber. The total annual costs for the three cases considered are plotted in Figure 11. Note that these costs do not include the clarifier costs.

Figure 11 shows that at short detention times, the annual cost per cubic foot increases in a curvilinear fashion. This is reasonable since the volumetric cost represents only a small portion of the annual costs and is relatively constant over the entire scale of detention times. However, the oxygen requirements are time dependent and as the detention time decreases, the rate at which oxygen must be supplied increases, resulting in the rapid increase in annual cost.

Another interesting point on the curves is the point at which the total cost for each aerobic reactor drops below volumetric plus mixing costs. At detention times greater than this point, the mixing requirements are greater than the oxygen requirements.

If the costs of the clarifier and sludge handling facilities for the aerobic reactor are now evaluated and added to the total cost of the aerobic reactor chamber, then a total annual cost for treating the waste aerobically can be determined.

According to the criteria being used in this evaluation, the standard surface loading rate for the clarifier is 800 gal/ft<sup>2</sup>/day for a MLVSS concentration of 2500 mg/L. Based only on the hydraulic loading rate, the required clarifier surface area is,

$$A_S = \frac{\text{influent flow rate}}{\text{overflow rate}} = \frac{1,000,000 \text{ gpd.}}{800 \text{ gal/ft}^2/\text{day}} .$$

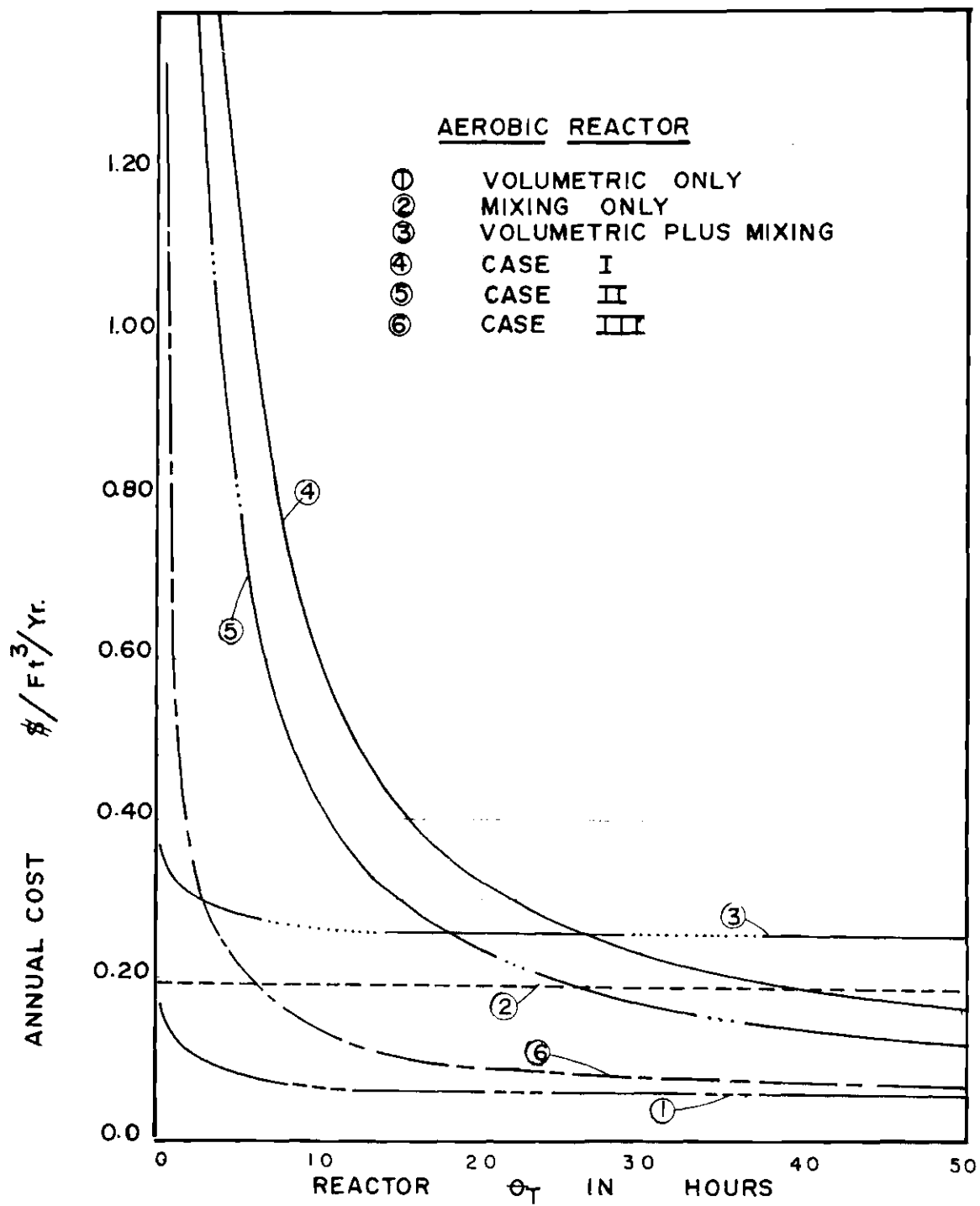


Figure 11. Annual Reactor Cost for Treating a Waste Aerobically as a Function of Detention Time.

However, the clarifier surface area is a function of both the hydraulic and solids loading, and therefore, the required surface area must be increased in direct proportion to the ratio of the MLVSS concentration to 2500 mg/L. Therefore, the required surface area is,

$$A_s = \frac{1,000,000}{(800) \frac{(2500)}{X_1^0}} = \frac{X_1^0}{2}, \quad (63)$$

where  $X_1^0$  is the reactor MLVSS concentration.

Since surface area is a function of the solids concentration entering the clarifier, the reactor MLVSS concentration for various detention times and organic loading must be determined. Equation (64) is the MLVSS concentration for the aerobic system.

$$X_1^0 = \frac{1}{k' \theta_T} \left[ \frac{X_0^N}{X_L^N} - 1 \right], \quad (64)$$

where  $k' = 0.02 \text{ Day}^{-1} (\text{mg MLVSS/L})^{-1}$ .

Determining  $X_1^0$  in Equation (64) for Case I ( $X_0^N = 1500 \text{ mg/L BOD}_5$  and  $X_1^N = 20 \text{ mg/L BOD}_5$ ) and Case III ( $X_0^N = 200 \text{ mg/L BOD}_5$  and  $X_1^N = 20 \text{ mg/L BOD}_5$ ) and substituting into Equation (63), one obtains the required clarifier surface area in square feet for each case as a function of the theoretical reactor hydraulic detention time.

Case I 
$$A_s = \frac{1850}{\theta_T}. \quad (65)$$



Case III

$$A_s = \frac{225}{\theta_T} . \quad (66)$$

From Equations (65) and (66), surface area for any detention time can be determined. The clarifier depth is set by the criteria, and thus the clarifier volume can be calculated. By using assumption *g*, the quantity of concrete required to enclose the clarifier volume can be calculated. The in-place cost of concrete is \$100.00 per cubic yard. Clarifier sludge collectors can be purchased for \$100.00 per foot of diameter, exclusive of motor and drive mechanism. The sludge collectors in the clarifiers require 2.4 horsepower drive motors.

The annual cost for the clarifiers represents both an initial investment (structure, sludge collectors, and drive motors) which must be amortized over the life of the facility and the operating cost (drive motors). Therefore, by calculating, at various detention times, the sum of the initial cost of the concrete for the clarifier plus the initial cost of the sludge collectors plus the initial cost of the drive motors and amortizing this sum over the ten-year period, the ACCR for the clarifier can be evaluated. The clarifiers require 2.4 horsepower drive motors which represents an annual operating cost of \$490.50 per year. The sum of the operating cost and ACCR for each detention time represents the total annual cost of the clarifier (Figure 12). If this total is added to the total annual cost of the aerobic reactor for each detention time for Case I and Case III, then the total annual cost of treating each waste aerobically is determined and is plotted in Figure 13.

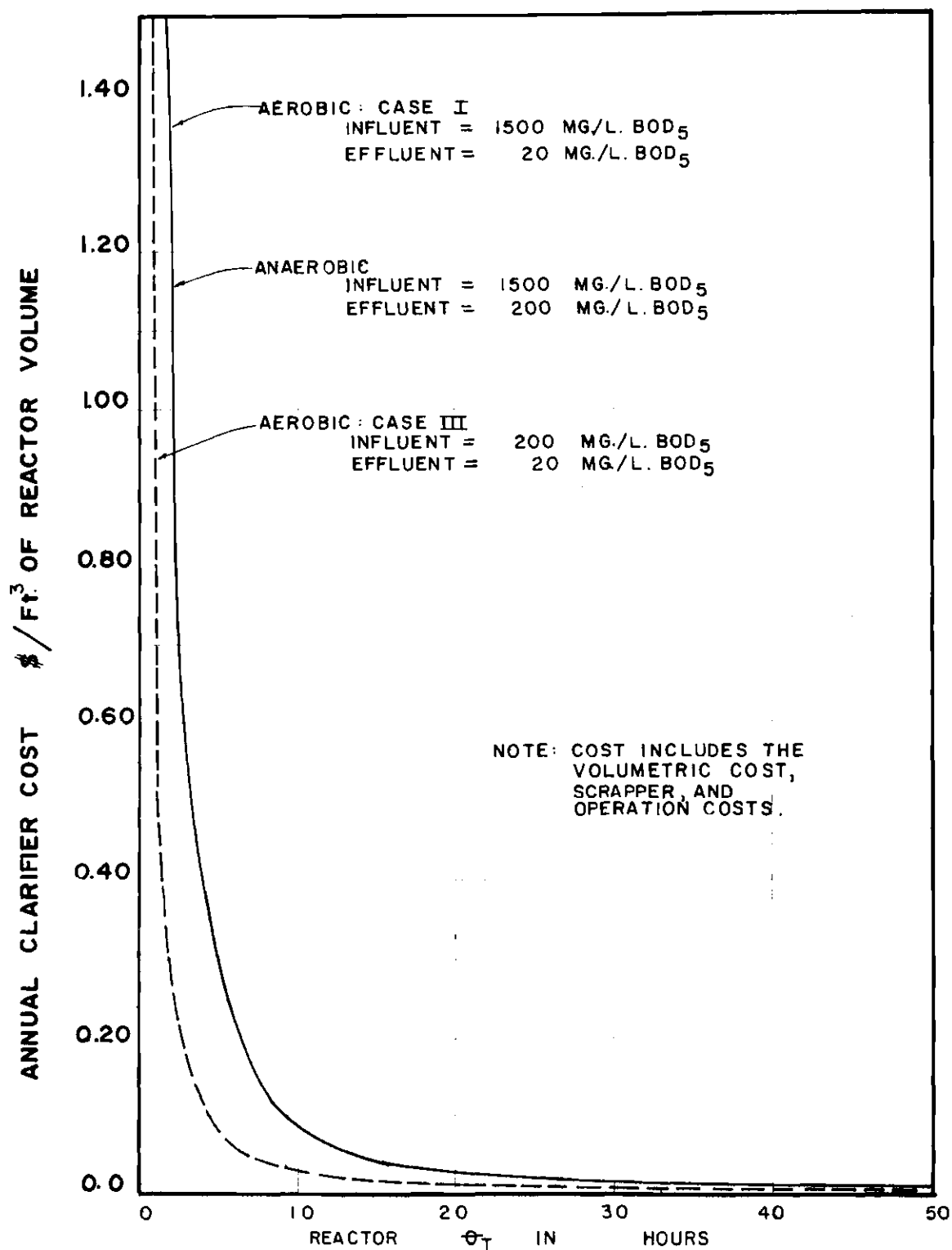


Figure 12. Annual Clarifier Cost for the Anaerobic Contact Process and the Activated Sludge Process as a Function of Detention Time.

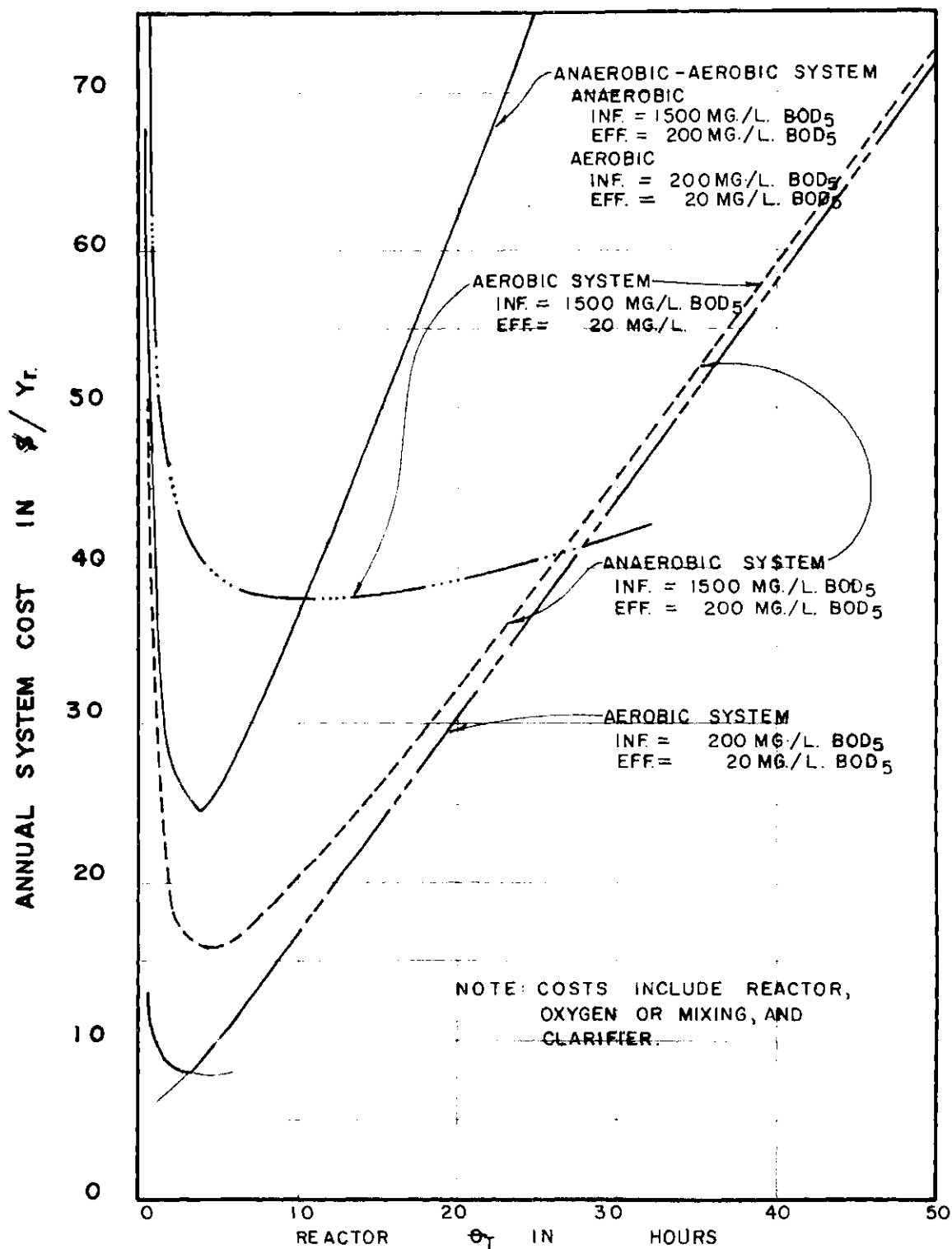


Figure 13. Total Annual Cost for Treating a Waste Anaerobically and Aerobically as a Function of Detention Time.

Note that the curve for Case I in Figure 13 changes slope instantly at a detention time of 26.5 hours. This is the point at which mixing requirements exceed the oxygen requirements. A similar point is apparent for Case III at a detention time of three hours.

#### Anaerobic Treatment Costs

An economic analysis of the anaerobic contact process is simplified since there is no oxygen requirement. The annual cost for treating a waste anaerobically will be limited to an initial investment for the reactor chamber, the clarifier, sludge collectors, and drive and mixing motors, and an operating cost for mixing and sludge collection.

The annual cost per cubic foot of reactor volume (including mixing) for anaerobic treatment has already been determined and plotted in Figure 10.

By evaluating the annual cost for the sludge clarifier and adding this sum to the annual cost for the reactor chamber, the total annual cost for treating a waste anaerobically will be determined.

The influent shall have a BOD<sub>5</sub> loading of 1500 mg/L and the effluent a BOD<sub>5</sub> value of 200 mg/L. Since the criteria limits the effluent quality to 20 mg/L BOD<sub>5</sub>, it is assumed that the anaerobic effluent will be passed through an aerobic polishing process to reduce the effluent BOD to the required 20 mg/L BOD<sub>5</sub>.

The required clarifier surface area can be found from Equation (63). The value of  $X_1^0$  for the anaerobic contact process is

$$X_1^0 = \frac{Y^0(X_0^N - X_1^N)}{k^D\theta_T + C^0} = \frac{Y^0(X_0^N - X_1^N)/\theta_T}{k^D + \frac{C^0}{\theta_T}}, \quad (67)$$

where:  $Y^O = 0.38$ .

$$k^D = 0.07 \text{ Day}^{-1}.$$

$$X_0^N = 1500 \text{ mg/L BOD}_5.$$

$$X_1^N = 200 \text{ mg/L BOD}_5.$$

The value of  $C^O/\theta_T$  can be calculated from

$$X_1^N = \frac{K(k^D\theta_T + C^O)}{k^m\theta_T - (k^D\theta_T + C^O)} = \frac{K(k^D + C^O/\theta_T)}{(k^m - k^D) - C^O/\theta_T}, \quad (68)$$

where:  $K = 24.3 \text{ mg/L}$ .

$$k^D = 0.07 \text{ Day}^{-1}.$$

$$k^m = 0.14 \text{ Day}^{-1}.$$

Solving Equation (68) for  $C^O/\theta_T$  yields a value of 0.0548. Substitution of this value into Equation (67) results in the MLVSS concentration for the anaerobic contact process as a function of detention time, or:

$$X_1^O = \frac{3955}{\theta_T}. \quad (69)$$

Substituting Equation (69) into Equation (63) yields the required clarifier surface area,

$$A_s = \frac{1978}{\theta_T}. \quad (70)$$

The surface area may then be utilized to determine the concrete requirements, sludge scrapers, and drive motors for the clarifier. An

analysis identical to that for the aerobic clarifier can be applied to determine the total annual clarifier cost (Figure 12). This cost, when added to the reactor cost (Figure 10), yields the total annual cost for treating the waste anaerobically. This total annual cost is plotted in Figure 13.

Since the effluent from the anaerobic process has a  $BOD_5$  value of 200 mg/L and the criteria requires a  $BOD_5$  value of 20 mg/L, it is necessary to polish the effluent with an aerobic process. Therefore, the aerobic cost of improving the anaerobic effluent to meet the criteria requirements is added to the cost of the anaerobic treatment (cost for Case III of the aerobic process added to the anaerobic process added to the anaerobic process). See Figure 13.

By this procedure, an economic comparison between an anaerobic-aerobic treatment system and an aerobic system can be made.

### Conclusion

The method developed in this chapter for the economic evaluation of a treatment system can be used as an example for an economic analysis of any treatment system. The common base for all treatment costs was the reactor detention time.

The volumetric and mixing unit costs for both the anaerobic and the aerobic systems were the same, since both were independent of treatment requirements. The clarifier cost should not have been the same since it is a function of the solids loading rate. This, in turn, is a function of the process parameters. By chance, in the example presented, the anaerobic and the aerobic (Case I) clarifier costs

happened to be the same.

Figure 13 clearly indicates that a certain range exists in which an anaerobic-aerobic treatment system may have a definite economic advantage over that of an aerobic system. In the example presented, at optimum operating conditions and for equivalent treatment, the anaerobic-aerobic system net an annual savings of \$13,000.00 over that of a conventional aerobic system. This fact emphasized one of the objectives of this thesis, i.e., that in order to treat certain waste waters for the least costs, treatment systems other than the conventional aerobic or anaerobic systems are needed. One such system is the anaerobic-aerobic system.

Furthermore, all the curves of Figure 13 had a very sharp and distinct "low point." The shape and relative position of the curves is dependent upon the values of the various biological growth rate parameters. Since the "low points" economically represent the optimum operating conditions, and since the cost values at these points are dependent upon the shape and position of the curves, it is imperative that one have accurate values of the biological growth rate constants. Slight variations in these constants may radically alter the position of the curves.

Evidence has been presented which supports the argument that an anaerobic-aerobic treatment system may provide the equivalent treatment of an aerobic unit and at less cost. Consequently, a laboratory size anaerobic-aerobic reactor was constructed to attempt to determine the feasibility of such a system and to attempt to substantiate the theoretical curves developed in Chapters II and III.

## CHAPTER V

### INSTRUMENTATION AND EQUIPMENT

An anaerobic-aerobic "bench" scale reactor was constructed for the purpose of determining the feasibility of successfully operating a combined anaerobic-aerobic unit and to conduct a laboratory evaluation of the theoretical performance equations and curves as previously developed. The reactor was comprised of four basic sections; (1) the anaerobic contact section, (2) the anaerobic sludge separation section, (3) the activated sludge section, and (4) the activated sludge separation section.

A rectangular column, four feet in depth, one foot square in inside cross-section, closed at the bottom, and open at the top, was constructed out of one-quarter inch plexi-glass; as were the other components of the reactor. This rectangular column provided the shell within which the four sections of the reactor were formed. The anaerobic separation unit, the anaerobic-aerobic separator, and the activated sludge separation unit were constructed as separate units which could be inserted into the rectangular shell to form the four sections of the reactor. Since the units could be removed and varied in position, variations in relative detention time in each unit were possible, as well as removal of the units for modifications and cleaning. The shell was equipped with sampling ports at two-inch intervals from top to bottom.



### The Anaerobic Section

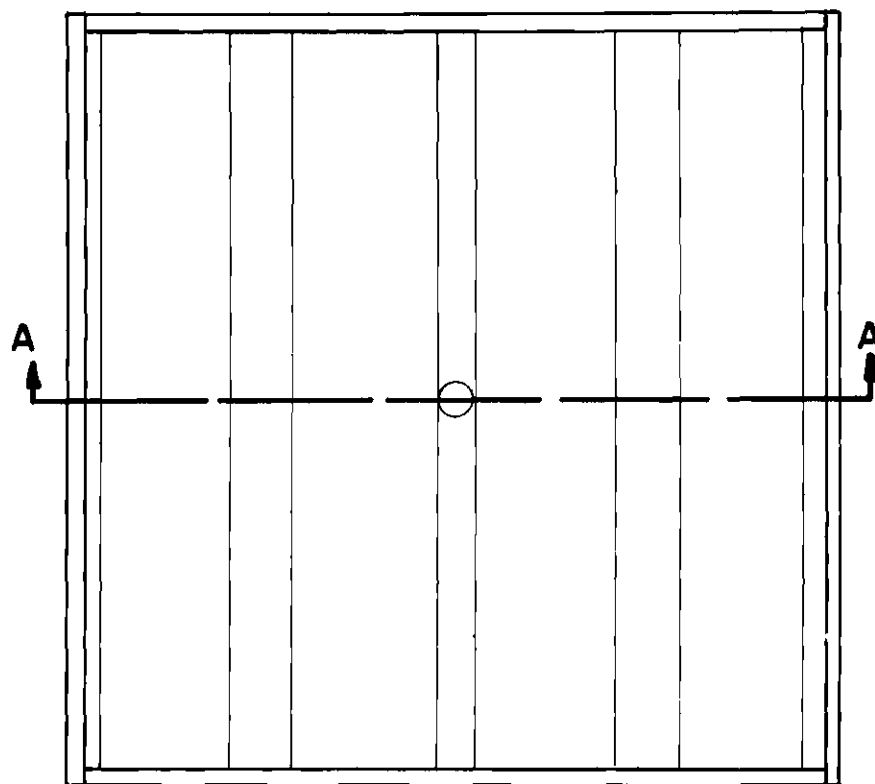
#### The Anaerobic Contact Section

The anaerobic contact section was in the bottom of the shell. The anaerobic sludge separation unit formed the top of this section. Consequently, this section was one foot square in cross section with a depth dependent on the position of the anaerobic sludge separation unit. Throughout all testing, however, the volume of this section was maintained at 27 liters. The section was mixed by a triangular-shaped plexiglass impeller on an aluminum shaft which extended the full depth of the reactor. The impeller had a face area of approximately 27 square inches and was turned at 20 rpm.

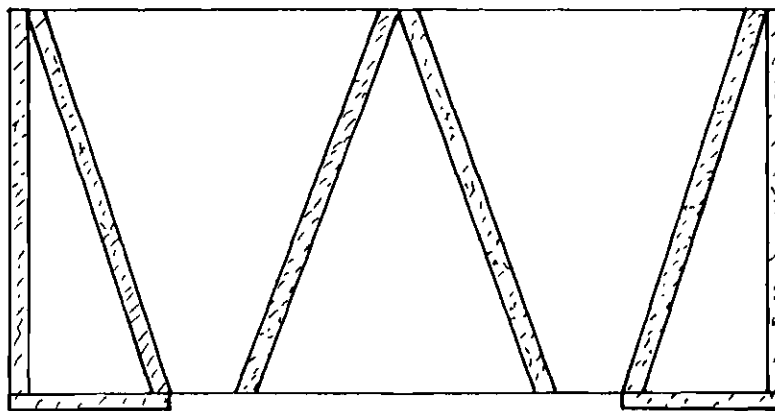
#### The Anaerobic Sludge Separation Unit

The anaerobic sludge separation unit consisted of two, trapezoidal-shaped upflow clarifiers having a combined area at the bottom of 0.159 square feet and a maximum combined area at the top of 0.839 square feet. The sides of the unit formed a square of one foot square outside dimensions so that a snug fit would be obtained between the sides and the shell wall.

Using a small surface area at the bottom, most of the turbulence created by mixing in the anaerobic contact section was not transferred into the separator. The turbulence which was carried over into the separator was dampened by the expanding cross sectional area. Details of the anaerobic sludge separation unit are given in Figure 14.



(A)



SECTION A-A

(B)

SCALE: 1" = 3"

Figure 14. (A). Top View of Anaerobic Sludge Separator.  
(B). Sectional View of Anaerobic Sludge Separator.

## Aerobic Section

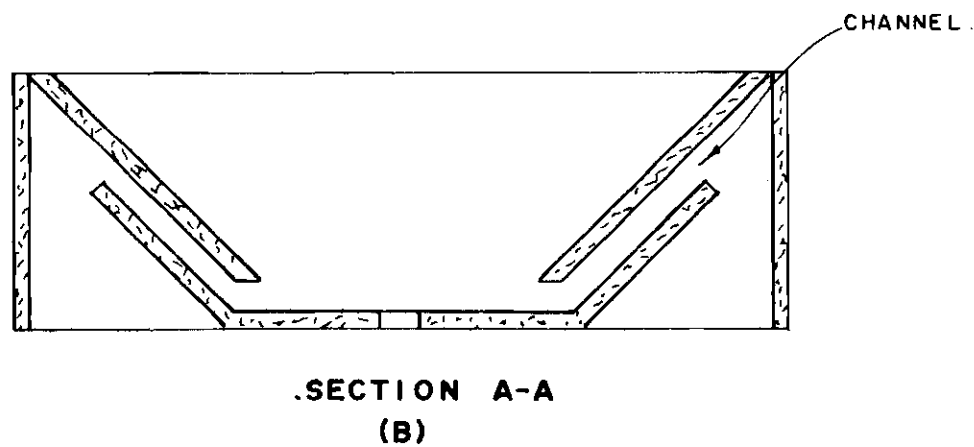
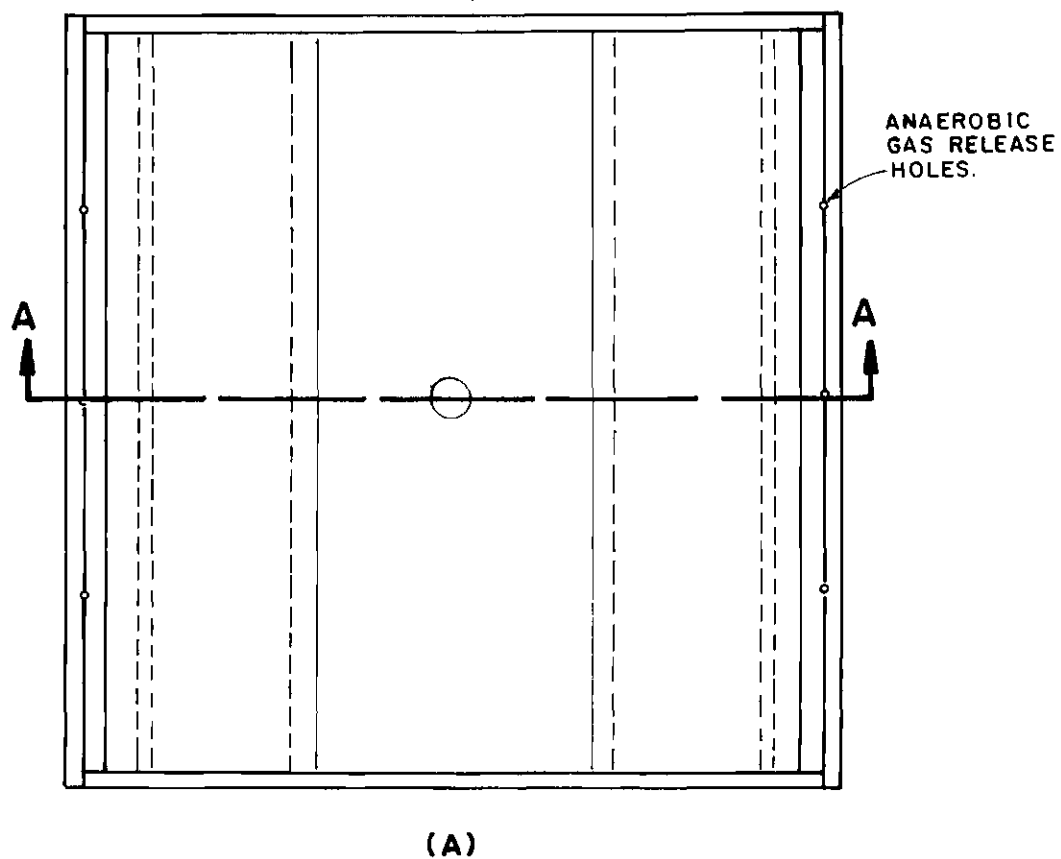
### The Anaerobic-Aerobic Separator

The anaerobic-aerobic separator was a partition between the activated sludge section and the anaerobic separation unit. Its objective was to prevent the flow of the activated sludge and oxygen into the anaerobic section and to provide the channels through which the anaerobic effluent entered the aerobic section.

The unit consisted of two channels on a 45° angle with the vertical. The combined cross sectional area perpendicular to the direction of flow was approximately 23 square inches. The anaerobic effluent was to pass upward into the mouth of the channels and then downward through the channels into the aerobic section. It was anticipated that the aerobic floc would be prevented from flowing through the channels into the anaerobic section by the turbulent pattern produced in the aerobic section and by the hydraulic loading on the channels from the anaerobic section.

The anaerobic-aerobic separator was provided with six, 1/8-inch diameter holes to permit the escape, into the aerobic section, of gas produced in the anaerobic section. The details of the separator as originally designed are presented in Figure 15.

However, due to difficulties encountered during the operation of the unit, several modifications of this section of the reactor had to be made. After several months of operation, aerobic sludge was discovered to be settling into the anaerobic section through the joint between the reactor shell and the anaerobic-aerobic separator and through the channels in the anaerobic-aerobic separator.



SCALE: 1"=3"

Figure 15. (A). Top View of the Anaerobic-Aerobic Separator as Originally Designed.  
(B). Sectional View of the Anaerobic-Aerobic Separator as Originally Designed.

Since the density of the mixed liquor in the aerobic section was greater than that of the anaerobic effluent and since the upflow velocity of anaerobic effluent into the aerobic section was small, causing the counter flow of sludge into the anaerobic section, it was necessary to completely seal off the aerobic section from the anaerobic section so that no flow passed through the channels or through the joint.

In order to provide for flow from the anaerobic section into the aerobic section but positively eliminate flow from the aerobic section into the anaerobic section, two 1.1-inch inside diameter plexiglass tubes were installed. These tubes passed through the bottom of the anaerobic-aerobic separator and into the anaerobic sludge separation unit. Since the flow in the reactor passed upward through the reactor, it could now pass up through these tubes into the aerobic section. To prevent counterflow of activated sludge down these tubes, the tubes were extended upward through the activated sludge section and above the liquid level maintained in the section. This permitted a free discharge into the section but eliminated any problem of counter flow. The final design of the anaerobic-aerobic separator is presented in Figure 16.

However, by sealing off the aerobic section at the anaerobic-aerobic separator, all openings for gas release from the anaerobic section were also sealed off. This necessitated further modification by the installation of two gas release ports through the reactor shell and into the triangular-shaped portion of the anaerobic-aerobic separator.

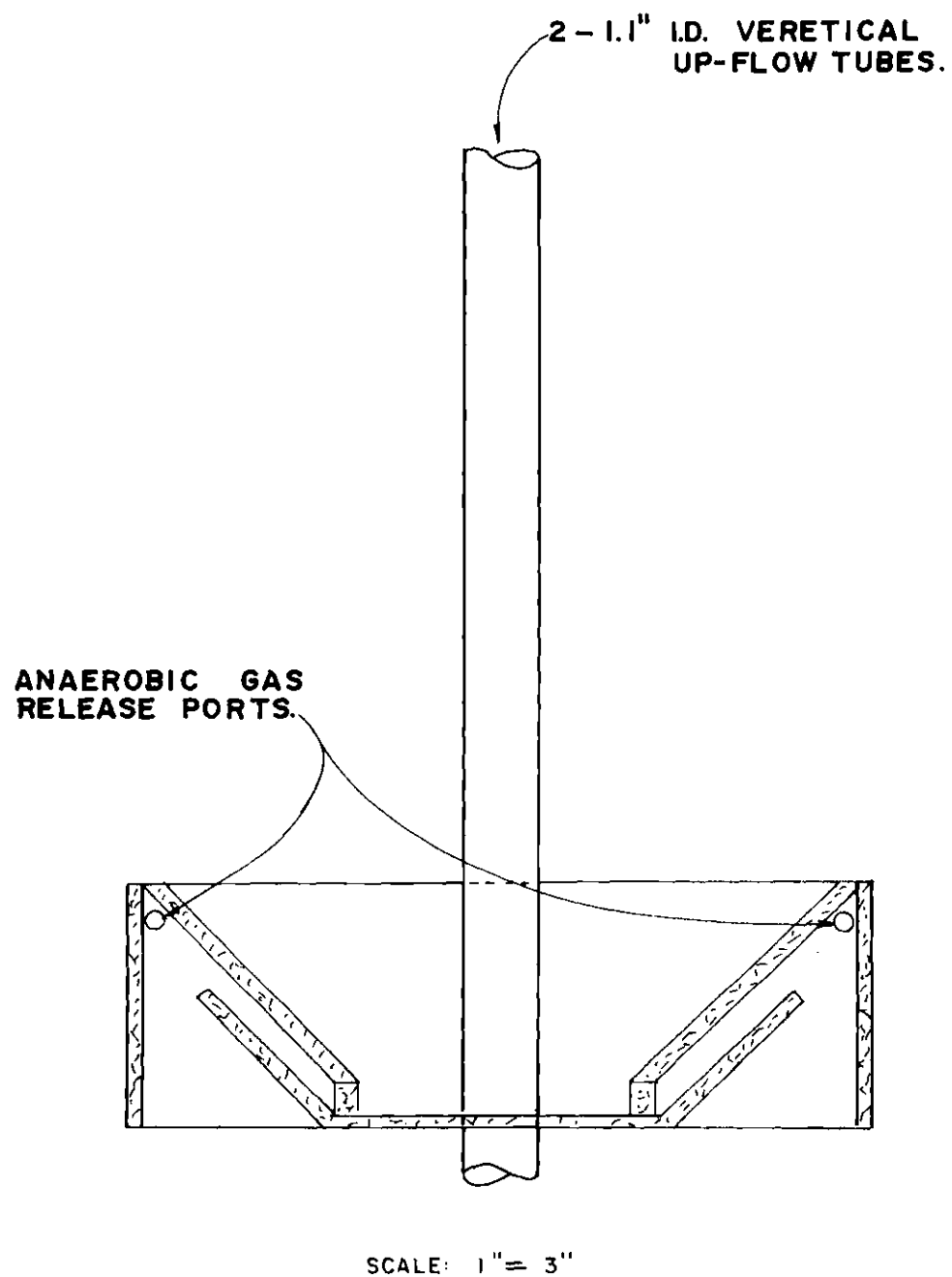


Figure 16. Sectional View of the Anaerobic-Aerobic Separator as Finally Designed.

Gas from the anaerobic digestion process would collect in these spaces and would require discharge either continuously or intermittently.

An efficient continuous flow gas release system was developed by connecting the gas release ports to flexible tubing which extended up the outside of the reactor shell and connected to one end of a glass rod. The other end of the rod was then submerged inside the reactor and into the bottom of the aerobic section. By proper adjustment of the air pressure in the aerobic section, the effluent from the anaerobic section would be forced under pressure up through the vertical tubes installed in the aerobic section rather than out through the gas release ports. This would permit a continuous collection and release of the anaerobic gases and prevent the gas release tubes from filling with liquid.

#### The Activated Sludge Separation Unit

The activated sludge separation unit was constructed in such a way that it formed both the clarifiers for the activated sludge and the top portion of the activated sludge section. The clarifiers were two, trapezoidal shaped upflow clarifiers having a combined bottom area of 0.079 square feet and a combined area at the top of 0.838 square feet. One side of each clarifier was vertical and the other at a  $30^\circ$  angle with the vertical. Whatever turbulence carried over from the activated sludge section would be dampened by the expanding cross sectional area and would aid flocculation in the clarifier. The sides at a  $30^\circ$  angle with the vertical would provide adequate slope for the settled biomass to slide down and back into the activated sludge section. The effluent from the clarifiers was discharged over two overflow weirs

having a combined length of 1.958 feet.

The air lines to the activated sludge section passed through the top of this unit, as did the air release line from the pressurized portion of the activated sludge section and the aluminum mixing shaft to the anaerobic contact section. A mercury seal was provided around the mixing shaft to maintain an air-tight seal for the aerobic section. The final design of the activated sludge separation unit is shown in Figures 17 and 18.

#### The Activated Sludge Section

The activated sludge section was the volume formed between the anaerobic-aerobic separator, the activated sludge separation unit, the pressurized air section, and the reactor housing. The volume of the section was dependent on the relative positions of the anaerobic-aerobic separator, the activated sludge separation unit, and the liquid level in the pressurized air section. It had a maximum volume of approximately 30 liters and a minimum volume of approximately 20 liters.

This section was completely mixed by diffused aeration through two carborundum diffusers supplied by a continuous air source. A circular mixing pattern was developed by the rising air bubbles in the center of the section, the 30° and 45° partitions of the activated sludge separation unit, and the anaerobic-aerobic separator. A pressurized air section was maintained in the upper portion of the section, just under the activated sludge separation unit, by submerging the air release line in a column of water. A schematic of the final design of the activated sludge section is presented in Figure 18. Figure 19 shows the reactor completely assembled.



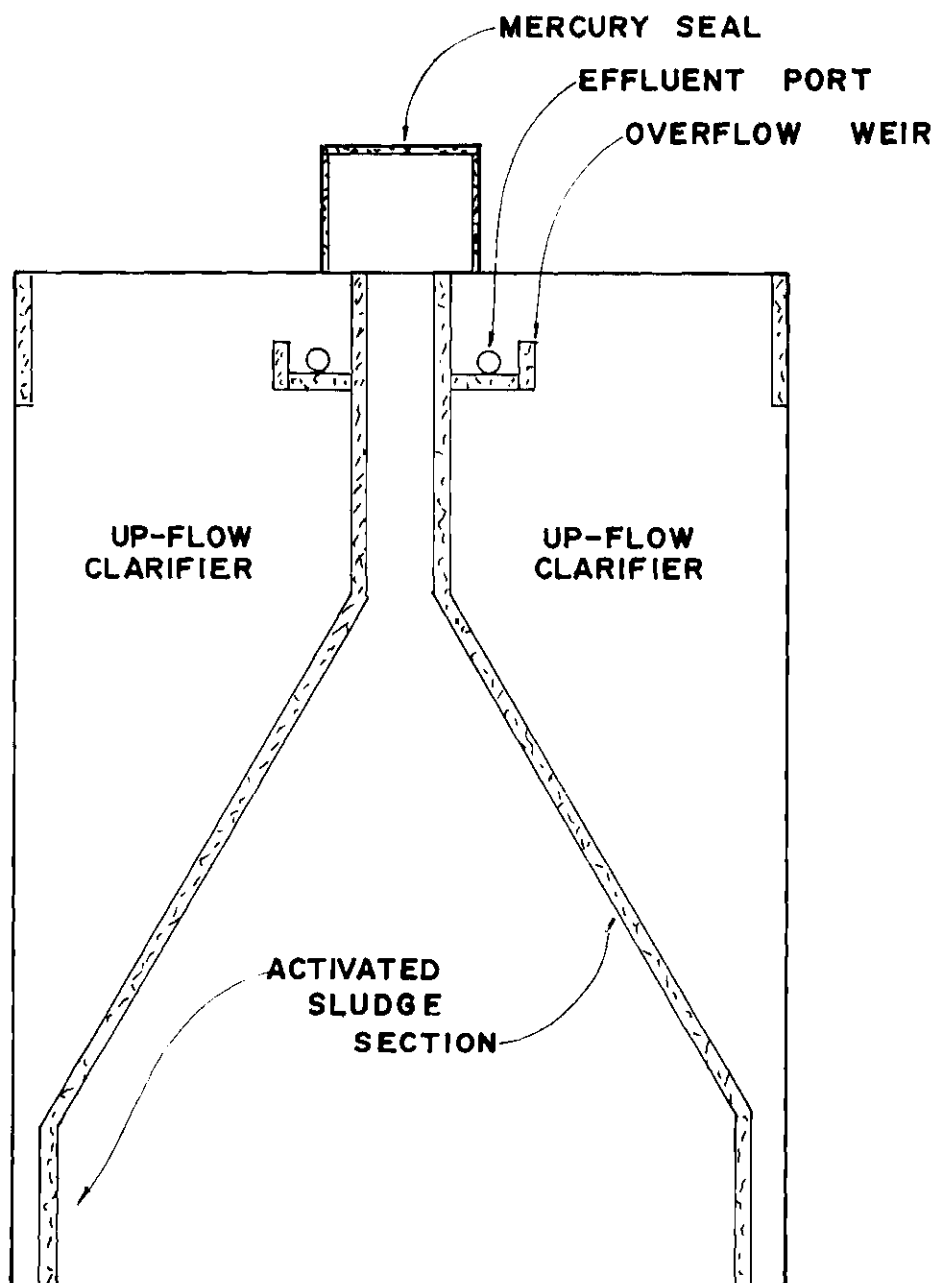


Figure 17. Sectional View of the Activated Sludge Unit as Finally Designed.

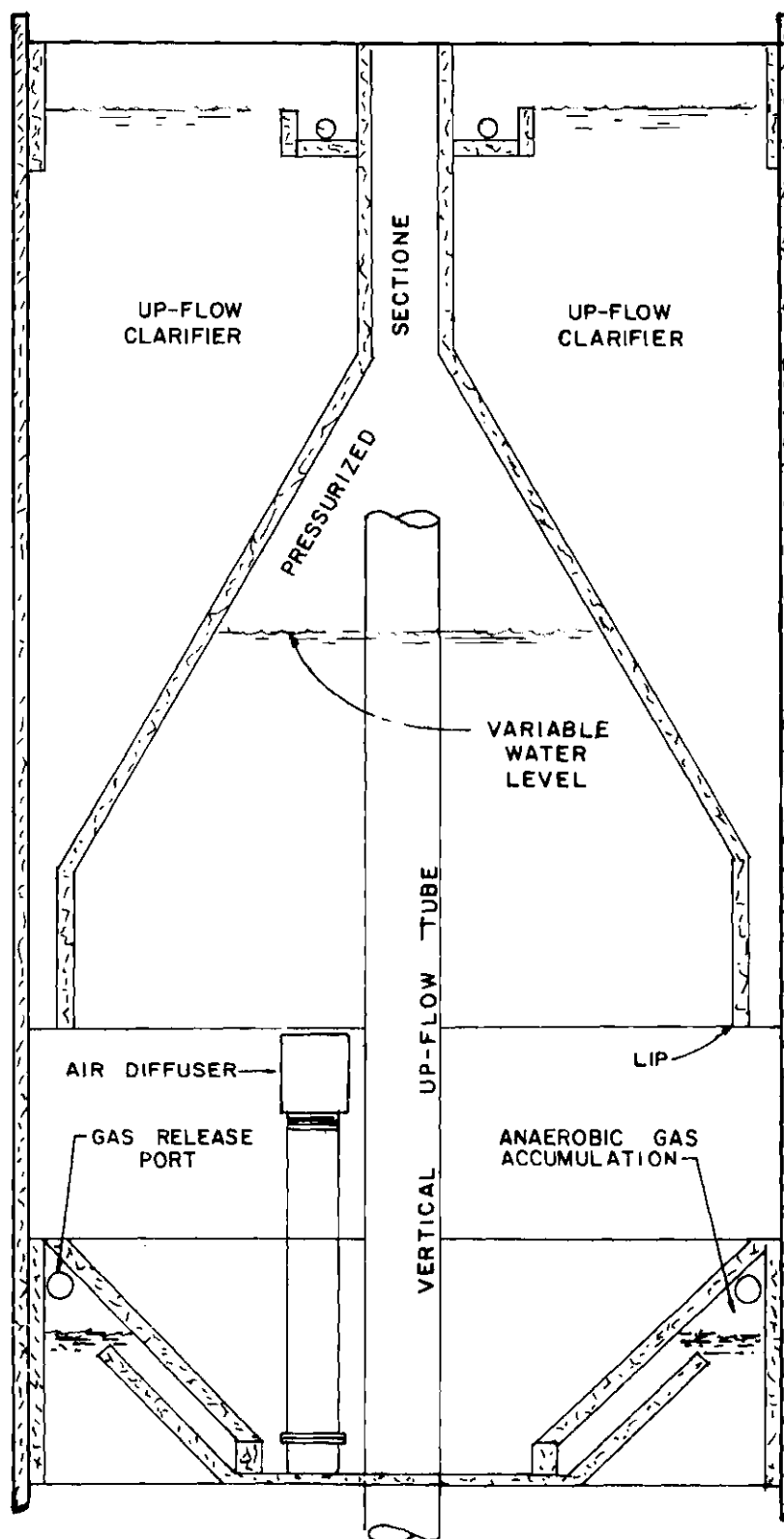
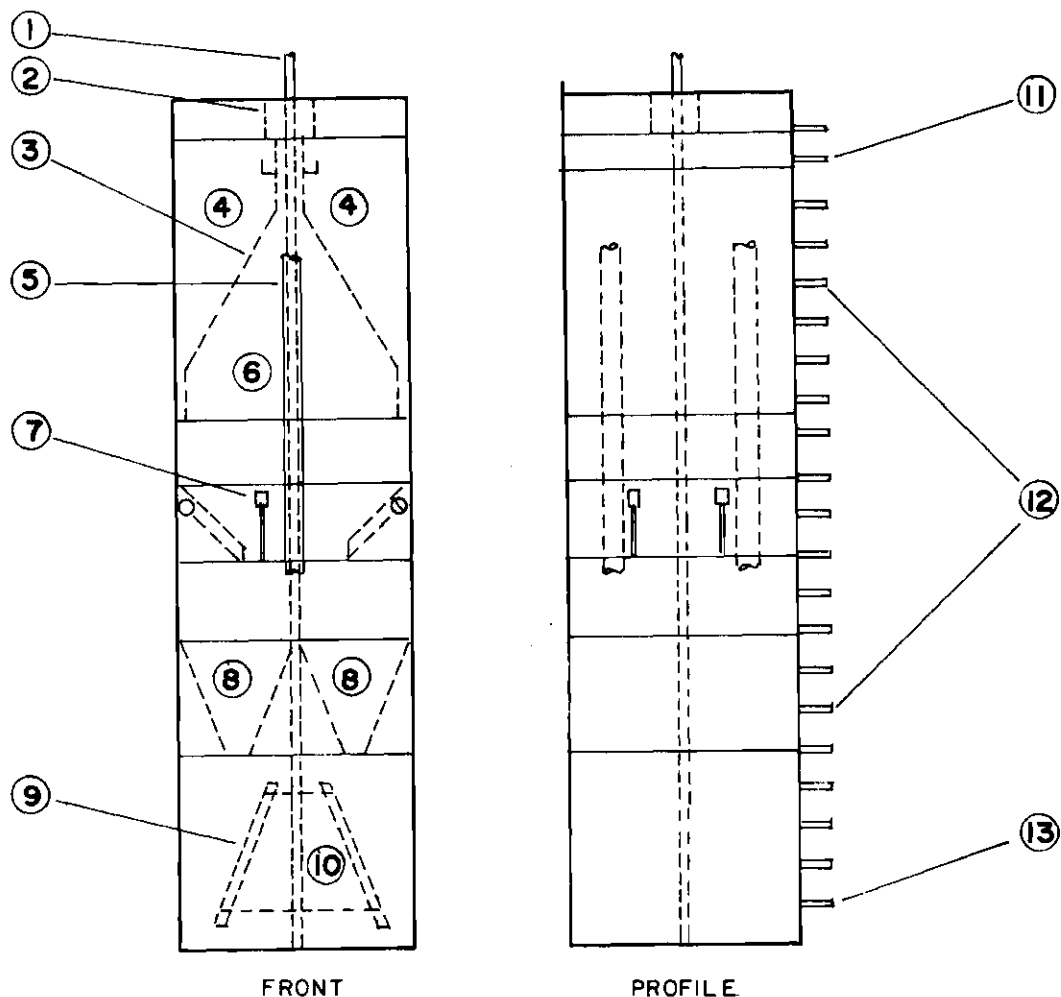


Figure 18. Schematic of the Activated Sludge Section as Finally Designed.



- ① ANAEROBIC CHAMBER MIXING SHAFT.
- ② MERCURY SEAL.
- ③ ACTIVATED SLUDGE SEPARATION UNIT.
- ④ AEROBIC UP-FLOW CLARIFIER.
- ⑤ VERTICAL UP-FLOW TUBE.
- ⑥ ACTIVATED SLUDGE SECTION.
- ⑦ AIR DIFFUSER.
- ⑧ ANAEROBIC SOLIDS SEPARATION UNIT.
- ⑨ ANAEROBIC CONTACT CHAMBER IMPELLER.
- ⑩ ANAEROBIC CONTACT SECTION.
- ⑪ EFFLUENT PORT.
- ⑫ SAMPLING PORTS.
- ⑬ INFLUENT PORT.

Figure 19. Complete Assembly of the Anaerobic—Aerobic Reactor as Finally Designed.

### Substrate Feed Tank

The substrate feed tank consisted of a plastic container calibrated to its 115 liter capacity. The substrate was mixed by a variable speed electric motor turning a one-half inch aluminum shaft with a 2-1/2 inch plexiglass impeller at sufficient speeds to prevent deposition of any solids. The base of the shaft rested in a steel thrust ball bearing held in a glass housing in the bottom of the feed tank.

### Pumping

A standard Sigmamotor T8 pump equipped with a zero-max Vernier speed control was used for pumping the substrate to the anaerobic chamber. This pump unit had a pumping capacity of zero to 14.4 liters per hour.

## CHAPTER VI

### PROCEDURE

#### The Apparatus

The various sections of the reactor were assembled in their relative positions within the shell. The volume of each section was calibrated by measuring the volume of clear tap water required to fill each section. In order to obtain an accurate volume measurement, the reactor was filled from the bottom-most sampling port. As each section was filled, the volume was noted. As a result, the anaerobic contact section had a volume of 27 liters, the anaerobic sludge separation unit a volume of 22 liters, the activated sludge section an approximate volume of 25 liters, and the activated sludge separation section an approximate volume of 18 liters.

The air diffusers were connected to the continuous air supply and adjusted to the maximum volume of air commensurate with satisfactory turbulence patterns.

The T8 Sigmamotor pump was elevated above the maximum reactor water elevation in order to prevent the reactor from being drained in the event that the pump cut a hole in the tubing. Tygon tubing having a 1/8-inch inside diameter and 1/16-inch walls was found to give satisfactory flow rates when pumping against the reactor's four feet of head.

### The Substrate Feed

The substrate was designed to simulate a 100 per cent soluble organic waste. The organic substrate utilized was Carnation instant nonfat dry milk. Sufficient nitrogen and phosphorus were added to the feed to bring the COD:N:P ratio to 100:25:1. Ammonium chloride was used for the nitrogen source. Initially, sodium phosphate was used to provide the phosphorus, but when it became difficult to maintain a satisfactory pH in the substrate chamber because of bacterial contamination, an equimolar concentration of monobasic and dibasic potassium phosphate at a 100th molar concentration was substituted for the sodium phosphate.

A fresh solution of substrate was prepared every 24 to 48 hours. The substrate tank was thoroughly washed with tap water at each new feeding. The tank was then refilled with tap water to the desired volume. A weighted amount of dry evaporated milk, ammonium chloride, and potassium phosphate was dissolved in the water and pumping and mixing resumed.

The pump operated 24 hours a day, 7 days a week (except for about 20 minutes each time a fresh feed solution was prepared). The substrate was pumped upward from the substrate tank, through the Sigmamotor pump, then down to the bottom-most sampling port, up through the anaerobic and aerobic chambers, and then out the effluent and into the effluent collection container.

### The Effluent Collection

The total effluent was collected daily in a 55-liter calibrated

glass container. It was anticipated that this would provide a more accurate means of measuring the actual flow rate than by sampling the flow intermittently. In addition, one would be able to observe the character of the total effluent on a daily basis.

#### Reactor Operation

The reactor was placed in operation by seeding the anaerobic contact section with anaerobic sludge from the primary digesters of the South River Municipal Treatment Plant of Atlanta, Georgia. The aerobic section was seeded with activated sludge from the same plant.

The activated sludge section was aerated by bubble aeration from the two carborundum air diffusers and placed under the desired air pressure by adjustment of the submergence depth of the air release line. The air flow rate was then adjusted to the maximum permissible.

The mixing in the anaerobic contact section was begun to provide a complete-mix system.

The substrate was prepared at the desired composition and pumping was begun at the previously determined flow rate. The effluent was continuously collected in the effluent container.

After the unit was initially placed in operation, it operated continuously, except for periods of mechanical or system failure, for seven months. However, during the first four months of operation, numerous mechanical and system failures were experienced. During this initial period of operation, the various parameters were never constant long enough to evaluate the system performance.

The mechanical failures were quickly resolved, but the problem

of maintaining a satisfactory biomass population continued to plague the operation of the unit. The sludge in both the anaerobic and aerobic section continuously disappeared.

Both the anaerobic and aerobic sections were seeded with sufficient sludge to produce zone sedimentation under quiescent conditions and, under normal operation, the sludge concentration in each section should have increased rather than decreased. Obviously, since the sludges disappeared, they were washed from the reactor and the washout rate exceeded the growth rate. The problem was to discover why.

Several mechanical failures undoubtedly contributed to the sludge washout and made it impossible to perform a material balance on each section. As a result of the workmanship in the construction of the reactor, the various sections did not fit flush with the shell walls. Consequently, there was a large amount of short circuiting in the anaerobic section between the anaerobic sludge separation unit and the reactor housing. The turbulence produced in the anaerobic contact section forced sludge up between the shell and the sludge separation unit and reduced the efficiency of the separation section.

In order to alleviate this problem, the reactor was disassembled and G. E. Silicon Construction Sealant used to seal the joints between the reactor shell and the anaerobic sludge separation unit.

Aerobic sludge was later discovered to be settling into the anaerobic section through the joint between the reactor shell and the anaerobic-aerobic separator and through the channels in the anaerobic-aerobic separator. Consequently, these passages also had to be sealed and other modifications made as explained on page 65.



However, the modifications did not completely correct the problem of sludge disappearance. Other attempts were made to discover the cause including:

- (1) a dilution BOD analysis on a sample from the aerobic section to determine if a toxicity problem existed. There was no evidence of one;
- (2) Nitrogen and phosphorus analysis on a sample from the aerobic section to determine if there were sufficient nutrients available. The results indicated 440 mg/L  $\text{NH}_4^+$ , 1500 mg/L  $\text{PO}_4^-$ , and 1000 mg/L COD;
- (3) The addition of organic and inorganic solids to both the aerobic and anaerobic sections to increase the density of the floc.

Finally after five difficult months of operation, both the anaerobic and aerobic sections were reseeded with sludges and for the following two months, the solids remained in the system and the unit operated fairly satisfactorily.

#### Chemical and Physical Analysis

Chemical Oxygen Demand (COD) and Solids Analysis were performed on the various sections of the reactor every 24 to 48 hours during the period of satisfactory operation.

For each set of analysis, the samples were collected as follows. Approximately 100 ml. of the influent was collected from a take-off port in the influent line immediately before it entered the anaerobic section. Approximately 250 ml samples were taken from the sampling

ports into the anaerobic contact section and the activated sludge section. A 100 ml. sample was taken from the anaerobic sludge separation section. A 100 ml. aerobic section effluent sample was taken from the unit's effluent line (not the effluent collection chamber). In all cases, the first portion of sample collected was discarded and all samples were immediately analyzed.

COD analyses were run on the filtered and unfiltered influent, the filtered and unfiltered anaerobic mixed liquor, the filtered anaerobic effluent (aerobic influent), the filtered aerobic mixed liquor, and the filtered and unfiltered aerobic effluent (unit effluent). To obtain the filtered samples, approximately 40 ml. of the mixed liquor was filtered through a glass fiber filter pad. For the analysis, 10 ml. of sample, 20 ml. of 0.25 N potassium dichromate, 10 ml. distilled water, and 40 ml. concentrated sulfuric acid were used. All samples were refluxed two hours and titrated with 0.25 N ferrous ammonium sulfate.

Total and volatile solids analysis were run on the anaerobic mixed liquor, anaerobic effluent, aerobic mixed liquor, and aerobic effluent. Glass fiber filter pads were used for the analysis. All samples were filtered to dryness but were not washed because of the difficulty of filtering such a high concentration of solids. Sample sizes ranged from 8.0 ml. to 40.0 ml. Duplicate samples were run on the anaerobic mixed liquor and the anaerobic mixed liquor and the aerobic mixed liquor in an attempt to obtain more reliable results. All samples remained in the 103°C oven for 24 hours and in the 600°C furnace for 30 minutes.

The pH of the influent, the anaerobic mixed liquor, the aerobic mixed liquor, and the aerobic effluent was checked during each sampling period.

All testing procedure was in accordance with *Standard Methods for the Examination of Water and Wastewater* (25).

#### Data

Because of the difficulties encountered during the period of operation, the data obtained throughout the investigations is inadequate and unreliable. Therefore, none of the data obtained will be presented. Since the system did become operational toward the end of the experimental period, it is emphasized that with additional research, the system can be evaluated.

## CHAPTER VII

### CONCLUSIONS AND RECOMMENDATIONS

#### Conclusions

1. The Monod Model has been shown to be applicable to the anaerobic contact process and to the activated sludge process.
2. Theoretical process performance equations and graphs have been developed to predict the performance of both the anaerobic contact process and the activated sludge process.
3. The two processes can be combined so that the performance of an anaerobic-aerobic process can be evaluated.
4. The anaerobic contact process has been shown to be capable, theoretically, of substantial BOD reductions at detention times as low as 0.25 days.
5. A method for economically evaluating the anaerobic and the aerobic processes has been developed.
6. Although the aerobic process is capable of an overall higher degree of treatment and at a faster rate than the anaerobic process, the anaerobic process has been shown to be the most economical for a particular situation.
7. Because of the sharp dip of the curves for the annual cost of treating a waste water (Figure 13), an accurate evaluation of the biological growth rate constants is essential.
8. A design for an anaerobic-aerobic treatment facility has been presented.

9. Although insufficient data was obtained to substantiate the theoretical process performance equations, the anaerobic-aerobic treatment unit was shown to be operationally feasible.

10. The anaerobic solids separator unit provided a maximum of 99 per cent solids recycle at a loading rate of 30 gal/ft<sup>2</sup>/day at the maximum cross section.

11. The aerobic solids separation unit provided a maximum of 98 per cent solids recycle at a loading rate of 150 gal/ft<sup>2</sup>/day at the minimum cross section.

12. The design of the anaerobic solids separation unit proved adequate, but several modifications of the activated sludge solids separation unit are needed.

13. Additional experimentation with an anaerobic-aerobic unit is required to obtain sufficient data to substantiate the theoretical equations and to further evaluate the biological growth rate constants.

#### Recommendations

It is recommended that further experimentation with an anaerobic-aerobic facility be conducted. Such investigations should be designed to evaluate both the biological growth rate constants for the two systems as well as the theoretical process performance equations. Since slight variations in the values of the growth rate constants radically alter the evaluation of the process performance equations and the economic evaluation, precise values of the constants must be obtained.

It is also recommended that the activated sludge solids separation unit should be redesigned. Since the minimum upflow velocity occurred at the top of the clarifier, the solids layer of greatest concentration also occurred at the top of the clarifier rather than at the bottom. Consequently, the clarifier should be redesigned to have the largest cross sectional area at the bottom and the smallest at the top. Furthermore, the air diffusers should be so located as to permit the maximum air flow rate desired, without producing such turbulence that the air bubbles are carried up under the lip of the solids separation unit. Such an occurrence results in solids flotation.

## APPENDIX

## APPENDIX A

## DEFINITION OF NOMENCLATURE

$E$	Enzyme
$S$	Substrate
$ES$	Enzyme-substrate complex
$P$	Reaction product
$x^E$	Total enzyme concentration in the system
$x^S$	Substrate concentration
$x^{ES}$	Enzyme-substrate complex concentration
$V$	Velocity of reaction
$V_{\max}$	Maximum velocity of reaction
$V$	Volume
$k$	Reaction rate constants
$k$	Biological growth rate constant (Monod's constant)
$k^m$	Maximum growth rate possible by the organism
$K$	Concentration of nutrient at which growth rate is half the maximum
$x^N$	Nutrient (organic) concentration
$x^O$	Organism concentration
$x^{ON}$	Organism-nutrient complex concentration
$y^O$	Gross organism yield $\left[ \frac{\text{mg. organism produced}}{\text{mg. nutrient removed}} \right]$



$Q$	Reactor influent flow rate
$X_0^N$	Influent nutrient concentration
$X_0^O$	Influent organism concentration
$X_1^N$	Reactor nutrient concentration
$X_1^O$	Reactor organism concentration (MLVSS)
$X_2^N$	Settler effluent nutrient concentration
$X_2^O$	Settler effluent organism concentration
$r$	Fraction of influent recycled
$r'$	Fraction of influent flow to waste
$C'$	Degree of concentration of organisms attained in solids separation unit ( $X_1^O$ underflow/ $X_0^O$ settler influent)
$C''$	Degree of concentration of nutrients attained in solids separation unit ( $X_1^N$ underflow/ $X_0^N$ settler influent)
$k^D$	Rate of endogenous respiration ( $\text{Day}^{-1}$ )
$C^O$	Recycle factor
$X_w^O$	Average solids loss from the settler
$\theta_T$	Theoretical detention time ( $V/Q$ )
$\theta_{Tc}$	Critical theoretical detention time
$k'$	Biological growth rate constant for the aerobic system

## APPENDIX B

Table B-1. Determination of Reactor Substrate Concentration ( $X_1^N$ ) for the Anaerobic Contact Process as a Function of Theoretical Detention Time ( $\theta_T$ ) for a  $C^0$  Value of 0.05

$\theta_T$ in Days	$1.701 \theta_T$ in mg/L	$24.3C^0$ in mg/L	$1.701 \theta_T + 24.3 C^0$ in mg/L	$0.07 \theta_T$	$0.07 \theta_T - C^0$	$\frac{1.701 \theta_T + 24.3 C^0}{0.07 \theta_T - C^0}$ in mg/L
.25	.425	1.215	1.640	.0175	-	-
.50	.850	1.215	2.065	.0350	-	-
.75	1.275	1.215	2.490	.0525	.0025	995.0
1.00	1.750	1.215	2.915	.0700	.0200	144.5
1.25	2.120	1.215	3.335	.0875	.0375	89.0
1.50	2.550	1.215	3.765	.1050	.0550	68.5
1.75	2.970	1.215	4.185	.1225	.0725	57.8
2.00	3.400	1.215	4.165	.1400	.0900	51.4
2.25	3.820	1.215	5.035	.1575	.1075	46.8
3.00	5.100	1.215	6.315	.2100	.1600	39.5

Where: $X_1^N = \frac{K(k^0\theta_T + C^0)}{K^m\theta_T - (k^0\theta_T + C^0)} = \frac{1.701 \theta_T + 24.3 C^0}{0.07 \theta_T - C^0}$
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Table B-2. Determination of Required Organism Concentration ( $X_1^0$ ) for the Anaerobic Contact Process as a Function of Theoretical Detention Time ( $\theta_T$ ) for a  $C^0$  Value of 0.05

$\theta_T$ in Days	$X_1^N$ in mg/L	$(X_0^N - X_1^N)$ in mg/L	$0.37 (X_0^N - X_1^N)$ in mg/L	$0.07 \theta_T + C^0$	$\frac{0.37 (X_0^N - X_1^N)}{0.07 \theta_T + C^0}$ in mg/L
.75	1000.0	0.0	0.0	.1025	0.0
.80	429.3	570.7	211.2	.1060	1992.4
.90	271.2	728.8	269.6	.1130	2385.8
1.00	145.5	854.5	316.2	.1200	2735.0
1.15	104.0	896.0	331.5	.1305	2540.2
1.25	89.0	911.0	337.1	.1375	2451.6
1.50	68.5	931.5	344.6	.1550	2223.2
1.75	57.8	942.2	349.6	.1725	2026.7
2.00	51.4	948.6	351.0	.1900	1847.4
2.25	46.8	953.2	352.8	.2075	1700.2
3.00	39.5	960.5	355.4	.2600	1366.9

Where:  $X_1^0 = \frac{Y^0(X_0^N - X_1^N)}{k \theta_T + C^0} = \frac{0.37(X_0^N - X_1^N)}{(0.07 \theta_T + C^0)}$

Table B-3. Determination of the Rate of Substrate Reduction for the Anaerobic Contact Process as a Function of Theoretical Detention Time for a  $C^0$  Value of 0.05

$\theta_T$ in Days	$X_1^N$ mg/L	$(X_0^N - X_1^N)$ mg/L	$\frac{X_0^N - X_1^N}{\theta_T}$ mg/L/Day
.75	995.0	5.0	6.67
.80	429.3	570.5	713.4
.90	271.2	728.8	809.8
1.00	145.5	854.5	854.5
1.15	104.0	896.0	779.1
1.25	89.0	911.0	728.8
1.50	68.5	931.5	621.0
1.75	57.8	942.2	538.4
2.25	46.8	953.2	423.6
3.00	39.5	960.5	320.2

## APPENDIX C

Table C-1. Determination of Reactor Substrate Concentration ( $X_1^N$ ) for the Activated Sludge Process as a Function of Theoretical Detention Time for a  $C^0$  Value of 0.50

$\theta_T$ in Days	4.167 in mg/L	$\frac{1}{k'Y^0\theta_T}$ in mg/L	$\frac{C^0}{k'Y^0\theta_T}$ in mg/L	$4.167 + \frac{C^0}{k'Y^0\theta_T}$ in mg/L
.25	4.167	333.33	167.0	171.0
.50	4.167	166.67	83.3	87.0
.75	4.167	111.11	55.5	59.5
1.00	4.167	83.33	41.7	45.7
1.25	4.167	66.67	33.3	37.3
1.50	4.167	41.67	20.8	24.8

Where:  $X_1^N = \frac{k^D\theta_T + C^0}{Y^0k'\theta_T} = 4.167 + \frac{C^0}{k'Y^0\theta_T}$

Table C-2. Determination of Required Organism Concentration ( $X_1^0$ ) for the Activated Sludge Process as a Function of Theoretical Detention Time for 90 Per Cent Removal Organic Loading

1	2	3	4	5	6	7	8
.25	4.0	200.00	.90	.10	10	9	1800
.50	2.0	100.00	.90	.10	10	9	900
.75	1.33	66.67	.90	.10	10	9	600
1.00	1.00	50.00	.90	.10	10	9	450
1.25	0.80	40.00	.90	.10	10	9	360
1.50	0.67	33.35	.90	.10	10	9	310
1.75	0.57	28.55	.90	.10	10	9	257
2.00	0.50	25.00	.90	.10	10	9	225

Where:

- 1 =  $\theta_T$  in Days
- 2 =  $1/\theta_T$  in Days<sup>-1</sup>
- 3 =  $1/k'\theta_T$  in mg/L
- 4 = Decimal fraction of BOD reduction
- 5 =  $1/(\text{Decimal fraction of BOD remaining})$
- 7 = Column #6 - 1.0
- 8 = (Column #3 times Column #7) =  $X_1^0$  in mg/L.

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